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for the refining-natural gasoline industry

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and The COMPOSITE CATALOG
for the drilling-producing-pipe line industry

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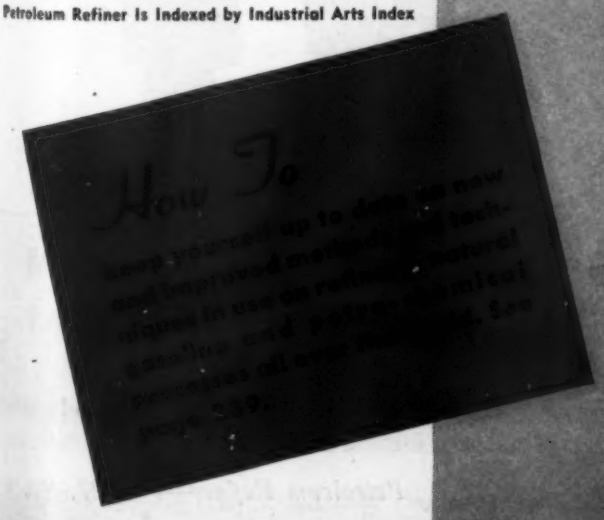

PETROLEUM REFINER

Specialized for the manufacturing division of the petroleum industry, refining, petrochemical processing, cycling and natural gasoline operations.

R. L. DUDLEY, President, Publisher A. L. BURNS, General Manager

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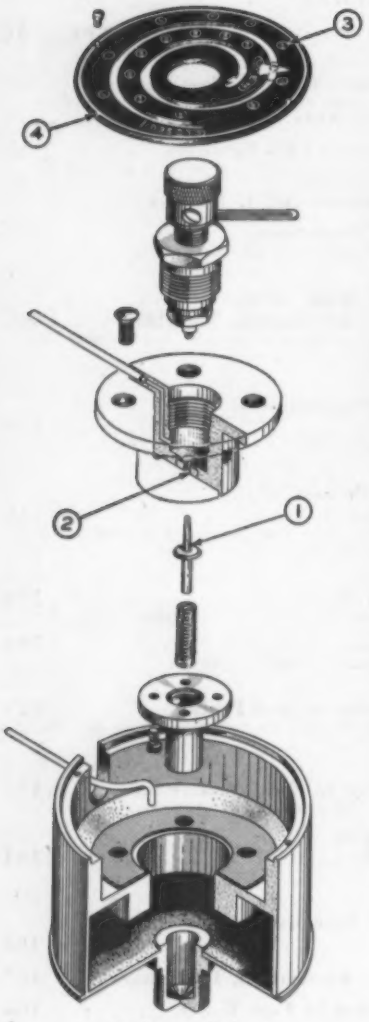

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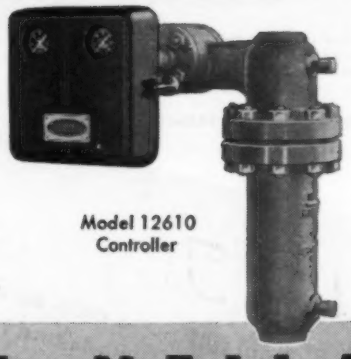

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VOLUME 27, NO. 4 APRIL, 1948

PETROLEUM
REFINER

SPECIALIZED FOR THE OIL REFINING INDUSTRY, INCLUDING PETROCHEMICAL ENGINEERING, PETROLEUM SYNTHETICS AND NATURAL GASOLINE

A GULF PUBLISHING COMPANY PUBLICATION

THE LOOK BOX

PARALLELING the annual technical meeting of the Natural Gasoline Association of America held in Fort Worth, Texas, on March 24, 25, and 26, it has seemed logical to devote this issue of PETROLEUM REFINER primarily to articles concerned with the design and operation of light hydrocarbon processing plants, or with the manufacture of petrochemicals from light hydrocarbons.

Special Number Memorandum

Also, since the independent activities of two technical groups have profoundly influenced the development of the industry to its present stable condition, it has seemed appropriate to include brief outlines of the histories and functions of both the Natural Gasoline Association of America and of the California Natural Gasoline Association. Both groups have worked long and steadily in such worthwhile projects as the standardization of test procedures for petroleum products and have rendered invaluable service to the industry as technical non-profit educational organizations.

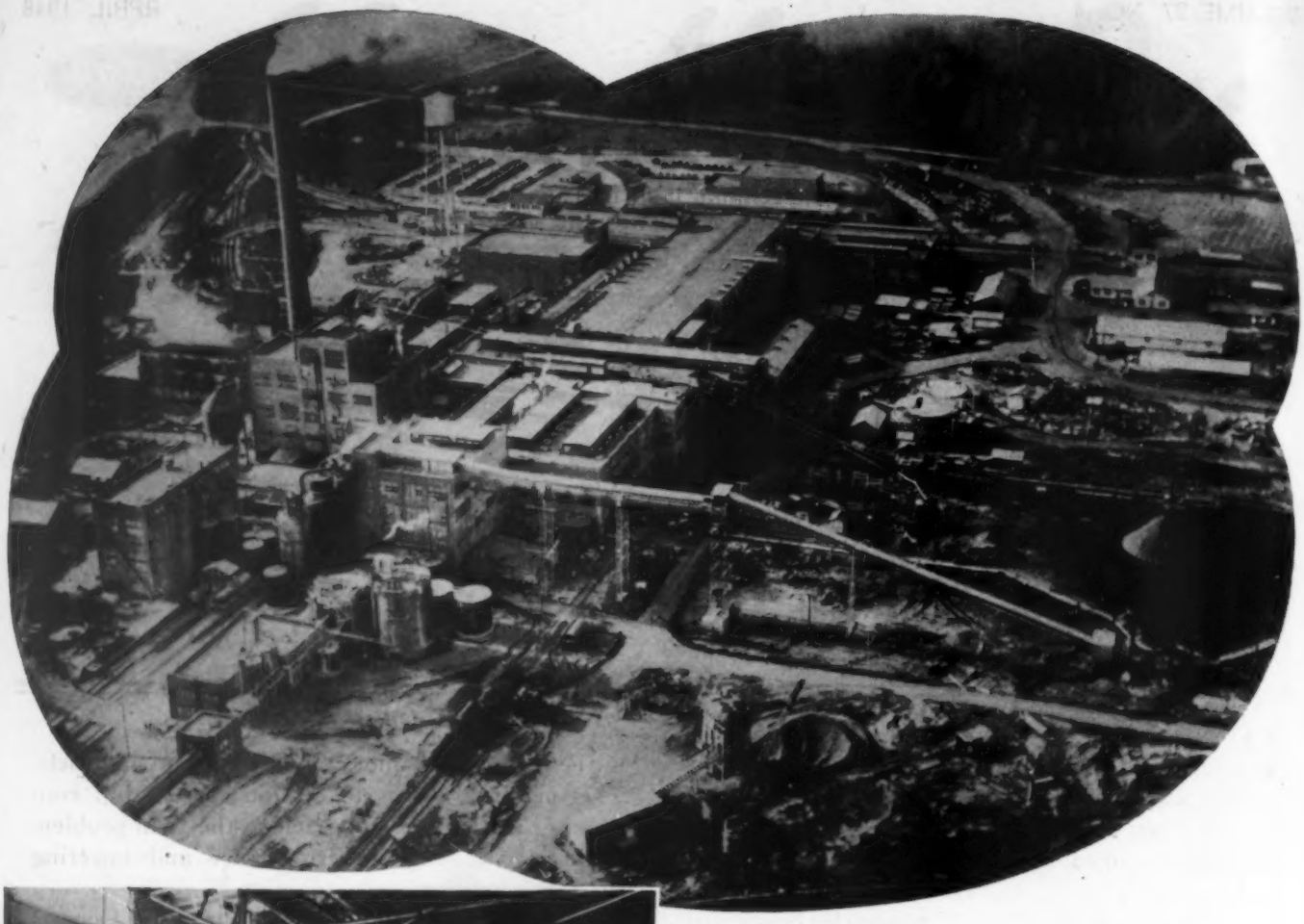
THE ever increasing demand for petroleum products will strongly influence and may alter the established routine of the past twenty-odd years by which the octane number requirements of automobile engines and the octane numbers of motor gasolines have followed a leap-frog path to the present high levels. At the recent NGAA meeting in Fort Worth, two speakers reported research work in the fields of automotive engine design and fuels. This material serves to outline current but contrasting methods of approach to the twin problem of improving engine performance and lowering fuel consumption.

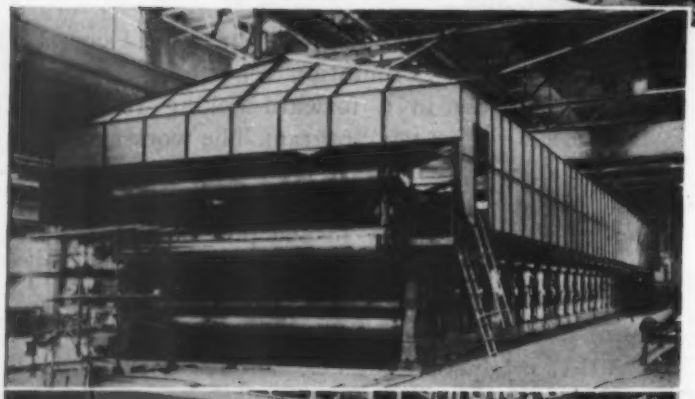
Altering the Leap Frog Path

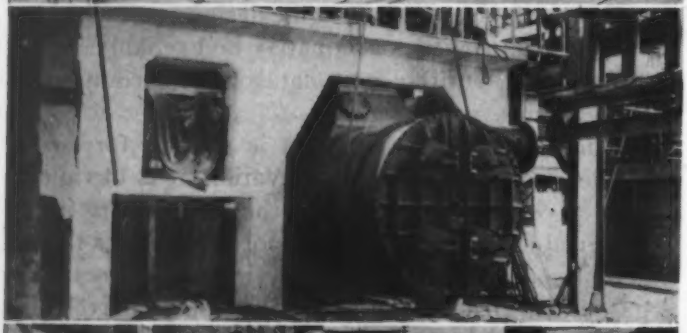
J. M. Campbell and L. L. Withrow, of General Motors Research Corporation, reported their company's progress in a forward-looking full-scale automobile engine test program. The work centers around a specially built engine of 12.5 compression ratio, requiring gasoline with Research octane number in the vicinity of 100 plus 0.2 ml. TEL and Motor octane number between 88 and 90. At constant engine performance, they report gains of 40 to 45 percent in fuel economy over corresponding 1947 production engines which required about 85 Research octane. The engine design, extensively road tested, performs adequately and reliably, and has supplied entirely new data on high compression engines.

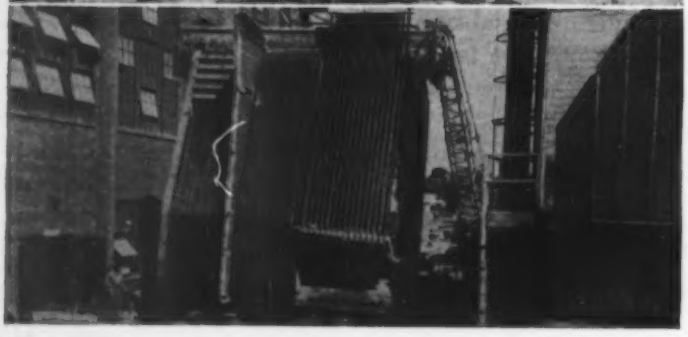
While this research project is a move forward in the direction of fuel conservation, it is obvious that a considerable part of the savings in fuel achieved by such engines is theoretical rather than actual conservation, since the yield of 100 octane gasoline which can be produced from a barrel of crude oil is substantially less than can be made at the 85 octane level.

Another approach to the same problems was discussed by C. H. Van Hartesveldt of Thompson Vita-Meter Corporation, who reported research on the injection of anti-detonants to improve automobile engine performance. Basic principle is that gasoline engines require high anti-knock fuel quality only when they are at or near full throttle.









how to multiply a paper mill by 2

At the Southland Paper Mills at Lufkin, Texas, above, Brown & Root, Inc., was given the problem of doubling the newsprint output.

175 Tons Daily X 2
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Doubling the production involved very substantial increases in power output, expansion of existing buildings, construction of new buildings, installation of new machinery, moving of old facilities, work of every type—and all the time manufacture of critical newsprint had to go on.

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Economics are reportedly favorable to the "two-fuel" system.

Distribution of anti-detonant liquid will constitute a major problem which is considerably simplified because the volume of detonant is normally much less than the corresponding gasoline fuel volume. This source of anti-knock quality is effective, of course, both to increase engine performance and to decrease fuel consumption. In practice it is entirely possible that extremely high compression engines, such as the 12.5 test engine mentioned above, can be satisfactorily operated on a two-fuel system consisting of a medium or even low octane gasoline and a suitable anti-detonant injection mixture.

In the final analysis, economics will control, but it is now evident that the old pattern of ever increasing octane numbers and process unit complexity is due for thoughtful evaluation.

Williams Brings NGAA Up to Date

ENCOURAGING facts and figures on their own industry have been heard repeatedly by natural gas-gasoline men since the time their operations began to assume the size and shape of a new industry. But disclosure of the information proving that it is a young, fully-grown, healthy industry with a bright future by C. R. Williams, president, Natural Gasoline Association of America at the March convention, brought expressions of satisfaction and amazement at the progress of the industry.

Among the many interesting facts that Williams pointed out to the assembly in his paper, "Significance of Liquids from Natural Gas," was the total production of liquids from natural gas—130 million barrels in 1947. Production, he said, of natural gasoline and heavier fractions increased 5.9 percent in 1947 above that of 1946, and liquefied petroleum gas production increased approximately 11 million barrels, an increase of 34.1 percent over 1946.

The production of natural gas since 1919 has increased almost five-fold, and the reserves of gas have increased ten-fold. Markets, he pointed out, are growing at an unprecedented rate. One of the most severe problems of the industry is distribution of the volatile materials to the consuming public due to the concentrated production of these materials in the nation's South Central section.

In presenting distribution percentages for 1947 over 1946, Williams showed that distribution of natural gasoline and condensate direct to jobbers and trade outlets increased 24.7 percent, LPG for fuel gas increased 43.6 percent. Of natural gasoline and heavier fractions distributed in 1947, approximately 83 percent was distributed to refineries for blending and further processing, and 15 percent sold directly to the consumer. The refiner received only 22 percent of the LPG production in 1947 from gasoline plants and of the total LPG produced, the refineries supplied 28 percent.

Domestic markets consumed 54.7 percent of the marketed production of LPG; chemical manufacturing, 22.5 percent; industrial uses, 15.5 percent, and gas manufacturing or enrichment, 7.3 percent.

Williams stated that the public was the largest consumer of LPG and the most difficult to serve, since this use and demand are seasonal and storage of summer production is required. It costs from 35 to 40 cents for steel storage of propane and from 20 to 25 cents for butanes. If the consumer would build his own storage, it would improve the industry's service to the consumer and would reduce the high capital investment required by the producer for storing his winter needs.

Gasoline produced from natural gas amounted to about 11 percent of the total motor fuel used in the nation in 1947. Williams also pointed out that this figure will undoubtedly increase in view of the 57 natural gasoline and cycling plants due to be completed in 1948 and 1949. These plants will produce an estimated 85,700 barrels daily additional gasoline, and 75,700 barrels daily additional LPG.

Completion of these plants will boost natural gasoline and LPG recovery to 59 million barrels additional per year, an increase of 45.4 percent over 1947. Of the 59 million, he said, 47 percent is for the recovery of additional LPG, and 53 percent is for the recovery of gasoline and heavier components. Texas will produce 79 percent and Louisiana 12 percent of the increased production.

Ethyl's Silver Anniversary

MORE than a billion barrels of oil have been saved for America since "Ethyl" gasoline was offered to the public 25 years ago, Charles F. Kettering, former General Motors research director estimated recently in marking the silver anniversary of the first sale of the new fuel.

During the 19-year period from 1929 through 1947 inclusive, when records of sales were kept, American motorists used 60.6 billion gallons of "Ethyl" gasoline, Ethyl Corporation reports.

In the same 19-year period, nearly 12 billion gallons of high-octane aviation gasoline were produced, and since 1933, the year the company started selling antiknock compounds for use in regular or house-brand gasolines, American motorists have used more than 250 billion gallons of such gasoline "containing tetraethyl lead."

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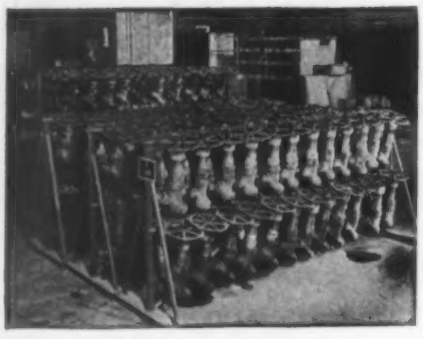


Ends capped, oiled and hung off the floor for protection against thread damage and dirt, these Edward forged steel valves are part of stock in an important Gulf coast supplier. In foreground, Edward 7500 lb Fig. 158 valves; in background, Fig. 2688.

Edward valves are stocked by distributors and oil field stores in all parts of the country. Some of the larger Edward valves are, of course, built on order, but all Edward forged steel valves, and many cast steel stop, check, non-return and other valve types are available out of stock.

If your jobber doesn't have the valve you want, he can get it for you and the chances are good that shipment can be made immediately.

Valve Values, Impactor and EValthrust are registered trade marks of Edward Valves, Inc.

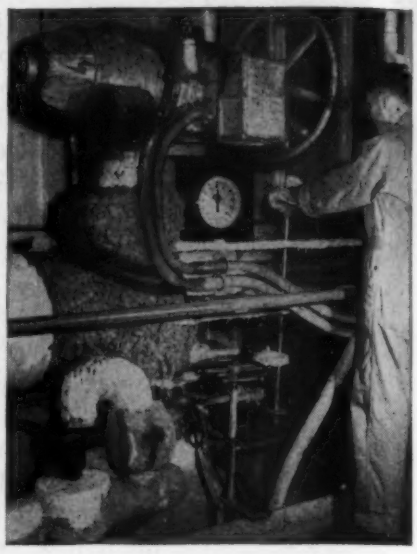
Pressure Drop Tests Take to Field

EDWARD LABORATORY FINDINGS CHECKED

Research into the characteristics of flow and the reduction of pressure drop has been a continuing Edward laboratory project for many years.

Recently Edward engineers checked their laboratory findings with a series of country-wide field tests, believed to be unparalleled in flow research, in large central stations operating under considerably different conditions.

Edward angle non-return valve as tapped for field test in 1300 lb 920 F midwestern central station.



Temporary insulation covers Edward motor-operated angle valve during in-service pressure drop tests

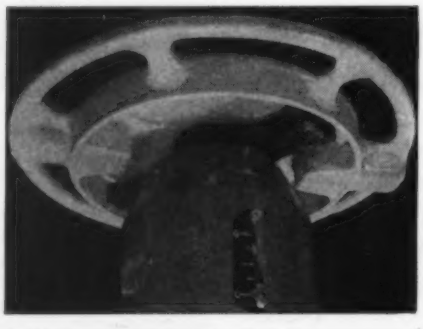
To make the tests, standard valves were drilled and tapped so that pressure recording gages could be installed at vital points in the path of flow. Stress analysis was made concurrently through strain gages affixed to the body of the valves.

Test results, coupled with laboratory data accumulated through the last decade, have guided the redesign of many types of Edward valves with reduction of pressure drop of up to 30 per cent.

IMPACTOR HANDWHEEL, CLOSURE INDICATOR AIDS TO OPERATOR

For years, the Edward Impactor handwheel has been the most popular method for manual operation of large valves. Usually attached to a valve with patented EValthrust ball bearing yoke, it has simplified closing big valves against high loads in close quarters.

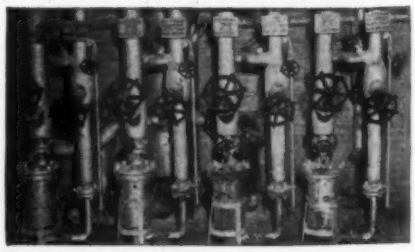
Now many large Edward valves are also equipped with a new, high-visibility closure indicator to show the position of the valve stem at a glance. In combination, the Impactor handwheel, EValthrust yoke bushing and Edward closure indicator are great helps to the operator.



Edward Fig. 2688's Used on Trap Lines

A good all-purpose forged steel globe valve is the Edward Fig. 2688, an O.S. & Y. valve, and its inside screw companion, Edward Fig. 2698.

A good example of the many piping uses for this valve series is this condensate and trap line hook-up on the



reheat system of a big new central station in north central United States.

The Edward inclined stem construction in small globe valves has many advantages, not only in lower pressure drop, but in reduced wear through freer flow, and, as the illustration shows, piping flexibility where a battery of valves must be installed in limited space.

EDWARD builds cast and forged steel globe and angle stop, gate, non-return, blow-off, feedline stop-check, INTEX (integral seat), integral bonnet UNIVALVES, relief, hydraulic, gage, straight-through valves and strainers for power, petroleum, marine, industrial and technological services.

Engineering Features of Hastings Plant

J. J. WEATHERBY and
HENRY N. WADE,
The Stearns-Roger Manufacturing Company

Distillation and Fractionation Structure, Absorbers, Stills and Fractionators, Cooling Tower at Stanolind Oil and Gas Company's Hastings, Texas, Gasoline Plant.

THE Hastings field gasoline plant recently completed by Stanolind Oil and Gas Company is located about 20 miles southeast of Houston in flat country with extremely poor drainage. On account of the poor bearing capacity of the ground, an unusually elaborate soil study was made as a preliminary to the design of the foundations. Cores were taken at equipment and building locations within the plant area, which totals about 40 acres. A complete set of compression tests was made on these cores. In addition, three consolidation tests were made from cores taken from two core holes bored to a depth of 45 feet.

Design of Foundations

The consolidation tests showed that a stratum having good bearing characteristics exists at depths varying generally between six and eight feet below the surface. It happened that this stratum dipped about two feet in the length of the compressor building, which is 411 feet. On account of this the foundation mat under the compressors was stepped, so as to take full advantage of the sound bearing stratum.

Elevation reference monuments were set in the bottom of core holes about 20 feet below the surface by the soil mechanics department of the University of Texas under the direction of Professor Raymond F. Dawson. The elevations of these monuments are transferred to the surface by means of ¾-inch straight steel pipes encased in 2-inch outer protective pipes, with the space between filled with oil. The top ends of these pipes are capped so as to prevent any possibility of water getting into them.

In order to determine the degree and rate of settling, bronze plugs were cast into each compressor block, with their rounded ends protruding slightly above the finished surface of the concrete. Similar plugs were set in various other critical foundations around the plant. The special detailed attention paid to the compressor foundations is due to the fact that the compressors represent the only application of reciprocating equipment around the plant, and it is expected that the vibration caused by their operation will eventually cause a much greater degree of settling than is likely to be experienced in other parts of the plant.

A careful level survey was made between all of the fixed elevation monuments, and the bronze plugs on foundation blocks were tied in to this survey as fast as the blocks were poured.

All foundations were designed on the basis of the recommendations of foundation consultants. It is interesting that, to date, the actual settling experienced has checked the predictions of the consultants within a few hundredths of an inch.

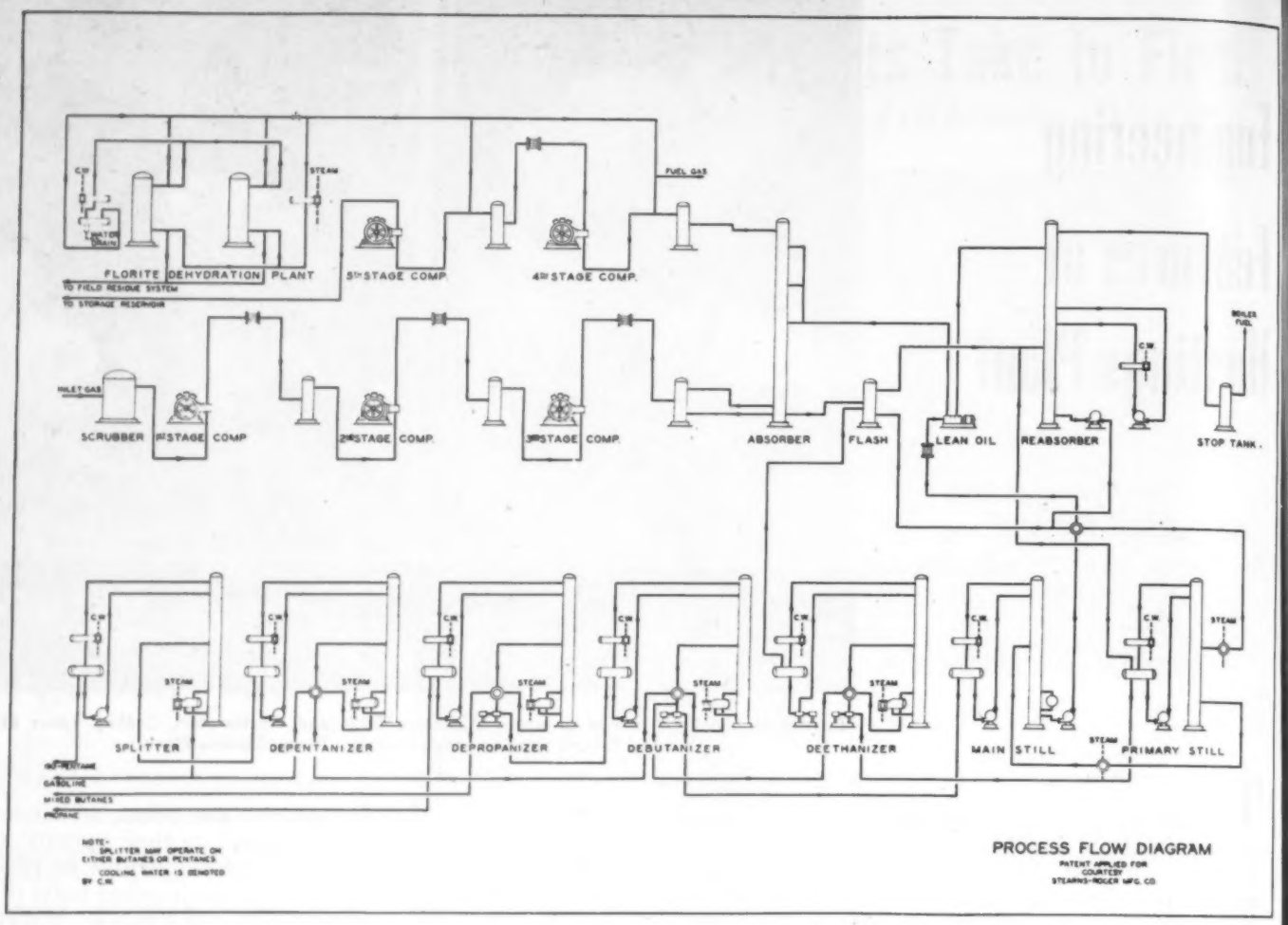
The plant is designed to handle 30 MMcf or below of gas per 24 hours, with full efficiency and a maxi-

IN THIS article, J. J. Weatherby, chief process engineer, and Henry N. Wade, consulting engineer, both of The Stearns-Roger Manufacturing Company, individually well known in the industry for progressive thought in the design of hydrocarbon processing plants, collaborate in a discussion of certain practical considerations that influenced the design of a modern gasoline plant.

The plant contains a number of engineering features that are somewhat out of the ordinary. It is to be noted that many decisions important in the design of the plant were made as a matter of judgment based on long experience.

Calculation procedures employed in the design of process plants are fairly well established in these days and workable precedents can usually be found for proposed processing schemes, so that engineers generally can turn out theoretically correct designs. If the resulting plants are to possess satisfactory operating characteristics, low maintenance requirements, high on-stream efficiencies, and sufficient flexibility to accommodate considerable change in products and feed specifications, something more than theoretical correctness is required.

Long experience and sound judgment are the most elusive elements of an engineer's necessary abilities, and are the ones least often assisted by the published literature. It thus seems likely that this material will hold high interest.



Process Flow Diagram of Stanolind's Hastings, Texas, Plant.

mum gas load of 40 MMcf per 24 hours with some loss in efficiency.

The gas is derived from 670 wells spread over about 5000 acres. Lead lines from the individual wells are brought in to a number of trap batteries where the oil and gas are separated and the gas is delivered at or somewhat above atmospheric pressure into the gasoline plant gas gathering system.

The gas is drawn into the plant by means of nine 1000-hp. angle type gas engine driven compressor units. The first stage suction cylinders on these units are sized so that they can handle the normal load of 30 MMcf at a suction pressure of about 10 psia. (10-inch mercury vacuum). By the use of clearance pockets the same compressors can handle the maximum load of 40 MMcf when the suction pressure is a few pounds above atmospheric.

The plant is designed to recover in finished products, each of not less than 95 percent purity where single components are called for, the following percentages of potential production contained in the wet gas: propane, 60 percent; iso-butane, 90 percent; normal butane, 95 percent; iso-pentane, 90 percent; normal pentane, and heavier 99-plus percent.

On the basis of present gas analysis this corresponds to the following quantities of production per day: propane, 360 barrels; iso-butane, 190 barrels; normal butane, 210 barrels; iso-pentane, 108 barrels, and debutanized gasoline 547 barrels.

Note that the separation of iso-butane from normal butane, or of iso-pentane from debutanized gasoline, is handled in a single fractionating column. On this account it is not possible to make iso-butane and iso-pentane as separate products simultaneously.

In order to attain the required high recovery of propane and butanes an elaborate study of absorber pressures was made so as to determine the most advantageous pressure. As a result of this study the absorber pressure of about 700 psig. was selected as giving the best compromise between oil circulation and secondary recovery requirements for vented gas.

Accordingly the field gas is compressed through three stages of compression to 700 psig. Gas cooling is provided together with efficient after-separation on the discharge from each stage of compression. The condensate from the compression and cooling is blended into the rich absorber oil.

Field fuel gas and gas lift gas, amounting to possibly 10 MMcf per day, may be taken directly from the outlet of the absorbers at 700 pounds. The rest of the gas from the absorbers is compressed through two stages by means of three 800-hp. angle type gas engine driven compressors to an intermediate pressure of 1550 psig. and an ultimate pressure ranging from about 3500 psig. up to a possible maximum of 5000 psig. In the event fuel and gas lift gas is required at higher than absorber pressure, these requirements can be supplied from the 1550 psig. gas.

The fuel and gas lift gas is cooled and dehydrated with a solid adsorbent type dehydration plant at either 1550 psig. or 700 psig. and then distributed back to the field.

The rest of the gas is compressed in the fifth stage to such pressure as is necessary to permit injection

into gas storage strata. At the present time this pressure is about 3000 psig. However, the cylinders and clearance pockets are designed so that the fifth stage can discharge at a maximum pressure of 5000 psig. The discharge gas from the fifth-stage compression is neither cooled nor dehydrated, but is delivered hot to the injection wells.

Since the pressure and volume requirements for gas lift are somewhat variable, the residue gas compressors are fitted with variable clearance arrangements so that the relative volumes and pressures delivered in the two stages may be varied over a considerable range. For example, if the fourth stage discharge pressure is 1550 psig., the fourth-stage cylinders can deliver the entire 700-pound residue gas, which, at a 30 MMcf wet gas load, amounts to 26.5 MMcf. If the fourth stage discharge pressure is raised to 1855 psig. this capacity is reduced to 21.9 MMcf. Similarly, the fifth stage with a suction pressure of 1550 psig. has a capacity of 18 MMcf. at a discharge pressure of 3500 psig. This same stage with a suction pressure of 1850 psig. has a capacity of only 14.9 MMcf. at 5000 psig. discharge pressure.

Seventeen hundred-pound Operating Pressure Florite Dehydrator for Drying Gas and for Gas Lift.

Absorption Oil Circuit

The wet gas compressed to 700 psig. is contacted with absorber oil in two 40-plate absorbers operating in parallel. The unusually large number of plates in the absorbers was selected to insure getting a maximum propane extraction with the oil circulation provided.

Since there has always been some question in the minds of plant designers as to whether there is any advantage in going much beyond 20 to 24 plates in an absorber, it was decided to arrange the Hastings absorbers so that the variation of extraction efficiency with varying number of plates could be investigated with all other variables held constant. In order to do this, additional oil input nozzles were provided over the thirtieth tray and over the twentieth tray, counting from the bottom. With this arrangement it is possible to compare directly the performance of one absorber against that of the other operating with a different number of trays, but with all other factors such as temperature, pressure, gas composition, oil ratio, etc., held exactly the same in both absorbers. So far only one complete set of comparative tests has been made.

The saturated absorber oil flows from the absorber to a flash tank which is held at a pressure of 250 psig. In this flash tank a large part of the methane dissolved in the oil at the higher pressure is released. Along with this methane an appreciable amount of propane and heavier components also weather out of the oil stream. In order to save these desirable components the gas from the flash tank is passed to a reabsorber which operates at 235 psig. This absorber is served with a separate stream of lean oil.

In order to increase the efficiency of the reabsorber a fat oil cooler is connected between a pair of plates near the bottom of the absorber. The entire stream of partly saturated oil is withdrawn from the absorber and is pumped through a closed cooler and back to the absorber.

The stream of fat oil from the reabsorber joins the main fat oil stream from the flash tank. The combined fat oil streams then pass through the main oil heat exchangers and through a steam heated preheater into the primary still.

The function of the primary or high-pressure still is to remove all methane and ethane, most of the propane, and part of the butanes and heavier, from the fat oil stream, leaving for removal in the main or low-pressure still only the heavier gasoline components which can be totally condensed at practically atmospheric pressure. Since the overhead product from the primary still is composed largely of pentane and lighter components, the de-ethanizer can work on it under easy conditions, especially with respect to base temperature requirements.

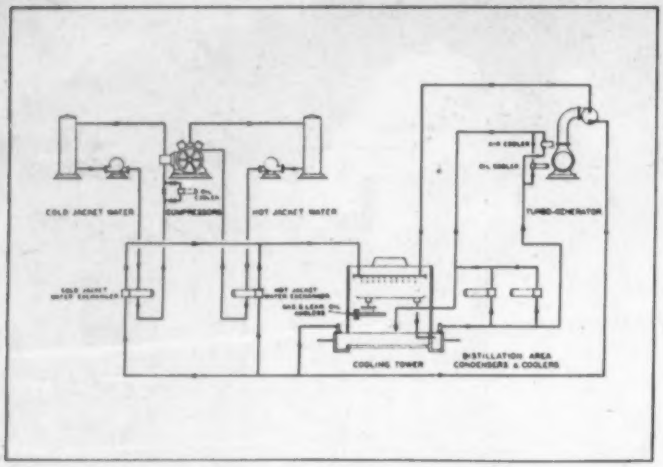
The partly stripped fat oil from the base of the primary still passes to the main still where the remaining heavy gasoline components are removed.

The completely stripped lean oil from the base of the main still is pumped through the lean side of the oil heat exchangers and through the final oil coolers to the suction of the high pressure lean oil pump, thus closing the absorption oil circuit.

The supply of lean absorption oil for the 235-pound reabsorber is obtained by taking a side stream at a suitable point from the multi-stage main lean oil pump.

Fractionation of Products

From the preceding description of the absorption oil circuit it is evident that all of the commercial products are separated from the fat oil stream from two separate sources. The first of these is a mixture of condensate and uncondensible vapor from the top of the primary still, and the second source is the to-



Cooling Water System.

tally condensed heavy gasoline from the main still.

In order to salvage as much as possible of the propane and heavier from the uncondensible vapors off the overhead product of the primary still, these vapors are returned to the base of the 235-pound reabsorber.

The condensible part of the overhead product from the primary still is pumped to the de-ethanizer which separates methane and ethane from the other components, with a minimum loss of propane in the overhead product. The uncondensible vapors from the overhead of the de-ethanizer are mingled with the fat absorber oil from the main high-pressure absorbers, so that the escaping propane is partially recovered through the vapors issuing from the 250-pound flash tank.

The base product from the de-ethanizer, consisting of propane and heavier components is mingled with the heavier gasoline from the main still, and the combined stream is fed to the debutanizer. Here the total stream of propane and heavier components is separated into two parts. The overhead stream consists of normal butane and lighter and the bottoms stream consists of iso-pentane and heavier.

The overhead stream from the debutanizer is then passed to the de-propanizer where commercially pure propane is separated as a top product, and mixed butanes are produced as a bottom product.

Depending on market demand, the mixed butanes are either run directly to storage or are passed to the splitter for division into pure cuts or normal and iso-butane.

The bottoms from the debutanizer are fed to the depentanizer where a top product consisting of all the iso-pentane and part of the normal pentane is separated from the main stream of heavy gasoline.

The stream of mixed pentanes from the top of the depentanizer may be passed to storage as such for subsequent reblending, or may be passed directly to the splitter for separation into a pure top product of iso-pentane and a bottom product consisting mainly of normal pentane. In this case the normal pentane is blended back into the heavy gasoline stream as indicated on the flow sheet.

Since only one splitter is provided for the separation either of pure iso-butane or of pure iso-pentane, it is obvious that both of these products cannot be made at the same time.

In normal operation, with the splitter operating on the mixed butanes, the plant produces as commercially pure separate cuts propane, iso-butane, normal butane, mixed pentanes, and the partially depentanized main stream of heavy natural gasoline. Normal butane and mixed pentanes can be reblended with the heavy gasoline to produce natural gasoline of any desired usual specifications. It is also possible to mix propane with surplus normal butane to produce LPG cuts of various specifications.

With the splitter operating on the pentanes the plant produces as separate commercially pure cuts propane, mixed iso- and normal butane, iso-pentane, normal pentane, and partially depentanized heavy natural gasoline. Various obvious combinations can also be made by reblending these products.

Steam System

Steam is generated in three integral furnace boilers each having a capacity of 30,000 pounds, per hour of steam at 400 psig. and 680° F. total temperature. Process steam throughout the plant is supplied either directly at full boiler pressure where required or indirectly at lower pressures by expansion through turbine drives, direct-acting steam pumps or pressure controlled make-up regulators.

The secondary steam systems operate at 90 psig. and 30 psig. The 400-pound steam system is used for the oil purifier preheater, the primary still preheater, and the main still reheater. Agitation steam for the primary still is throttled from the 400-pound system. The 90-pound system supplies the debutanizer and de-ethanizer reboilers while the 30-pound system supplies agitation for the main still and also heats the reboilers for the depropanizer, depentanizer and splitter.

All process and power steam is eventually condensed and the condensed water is salvaged by passing it through a large skimmer tank and a filter set.

All raw water make-up for the boilers is obtained by passing the raw water first through a deaerating heater and then to a pressure feedwater evaporator. The distilled make-up water is then blended with the clean condensed water returned from the general plant system, and the total stream is again passed through a deaerating heater before being fed to the boilers.

Power System

Electric power for the operation of pumps, cooling tower fans, lights and miscellaneous services about the plant is generated by three 750-kw condensing turbo generators. Power is generated at 440 volts for general distribution about the plant. Any two of these units will provide ample power for all plant operations.

In order to provide lights in the boiler house and power plant during an emergency shutdown at night an auxiliary gas engine driven unit is provided in the boiler house.

The electrical distribution system is divided into two parallel bus and feeder systems which extend to all important services throughout the plant. In the event of serious trouble on one of these systems it can be isolated and the plant can continue operation at somewhat reduced capacity on the other bus and feeder system.

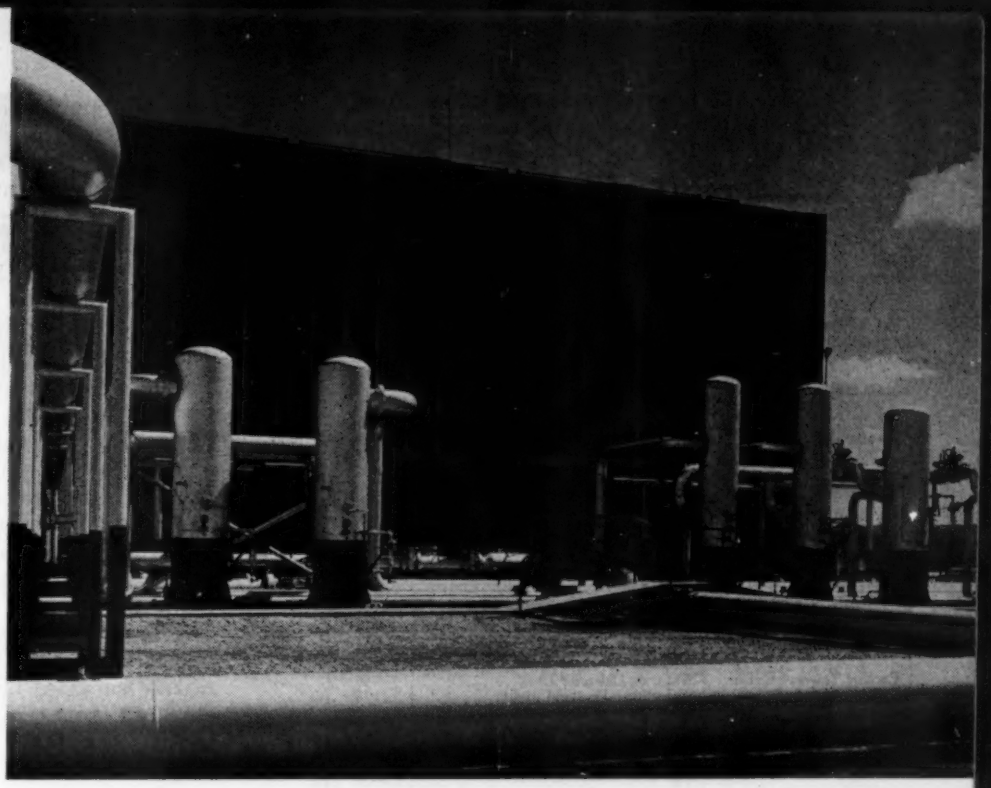
Cooling Water System

All plant cooling is accomplished by circulating water over a fan induced draft type cooling tower. As indicated on the cooling water flow sheet, heated water from all sources is delivered to the top of the tower where it is cooled down to within about 5

degrees above wet bulb temperature (under summer conditions) by the time it reaches the collector deck toward the bottom of the tower.

Here part of the water passes to an open spray system in the basin of the tower for distribution over the open cooling sections used for the gas and lean oil cooling. The rest of the cold water from the collector deck passes down into a closed sump just outside the tower basin from which it is pumped through closed condensers and coolers in the plant distillation area and also through the oil and air coolers for the turbo-generator sets. The hot water from this branch of the cooling system then returns to the main cooling tower basin.

Water from the cooling tower basin, warmed partially by passing through the condenser circuit just described and also by contact with the open coolers in the tower, is pumped through closed exchangers which remove the heat from the two seperate jacket water circuits associated with the compressors. One of these circuits serves the engine jackets only and operates at a comparatively high temperature level. The other circuit serves the engine oil coolers and compressor cylinder jackets.



North Side of Cooling Tower, Showing Dual Scrubbers for Interstage Cooling of Split Streams of Wet Gas.

Part of the basin water stream also goes through the turbo-generator condensers. The combined basin water streams are then returned over the top of the tower, closing the cooling water cycle.

By this arrangement a large part of the cooling water is passed through two groups of apparatus in series, supplying low temperature cooling where needed and moderate temperature cooling at points where close approaches to wet bulb are not necessary. This general arrangement effects a large saving in total cooling water circulation over the cooling tower.

Miscellaneous Unusual Features

Split Main Gas Flow. On account of the need to maintain uninterrupted field fuel and gas lift supply, the plant is arranged so that the incoming gas is divided into two separate parallel streams all the way from the main suction headers through to the outlets from the two 700-pound main absorbers.

The field gas enters the plant through two separate scrubbers which feed the first stage suction header at opposite ends. Block valves are provided in this header, as well as in all others, so that the compressors handling the first three stages can be divided into two groups. Separate suction and discharge lines are run from the headers serving each of these groups back and forth to the cooling tower, and the interstage gas coolers and scrubbers are in two separate parallel series.

Normally the plant is operated with the header block valves opened so that in effect there are simply two parallel gas streams working through the plant. However, by closing one set of block valves it is possible to isolate all of the gas piping and compressors on one-half of the system from the low pressure plant can be maintained at something like 50 percent suction up to the outlet of one of the high-pressure absorbers. In this way operation of the rest of the of capacity while extensive inspection and repairs are carried out on the other half.

Arrangement of Distillation and Fractionation Equipment. Since all condensers on the primary and main stills and on all fractionators are of the tube-bundle-and-shell type, it was decided to follow more or less conventional refinery practice in the arrangement of condensers, accumulators and pumps relative to the columns concerned. It is believed that this is the first time this sort of design has been applied to such a complete degree in a gasoline plant.

The stills and fractionating columns are arranged in a single row at one side of the distillation service structure. The distillation service structure consists of a three-story steel frame carrying at its top the condensers and aftercoolers for the distillation and fractionation columns. The accumulators for condensate from the condensers are mounted immediately beneath their respective condensers on the second deck of the structure. Finally, the reflux and other service pumps are mounted on the ground floor directly beneath the respective accumulator tanks.

Through this arrangement the connections from the various columns to their condensers, from the condensers to the accumulators, and from the accumulators to the pumps are made unusually direct and short, with consequent elimination of time lag and surging.

One of the results of this arrangement is exceptionally smooth operation of all of the column controls, which in turn has proved to produce unusual degrees of purity in the fractionated products.

Since it was decided to house the main high-pressure oil pumps and the control instruments along with the rest of the pumps in the base of the structure, it was necessary to extend the structure for the bottom story for a greater length than was required

TYPICAL WINTER OPERATING CONDITIONS
Average: Dry Bulb, 48°F., Wet Bulb, 40°F.

APPARATUS	Flow	Temp. °F.	Press. psig
Main Absorbers	Gas In	59	685
	Gas Out	73	....
	Oil In	69	....
	Oil Out	80	....
Reabsorber	Gas Out	64	210
	Oil In	69	....
	Oil Out	87	....
Fat Oil Flash	Gas Out	77	250
Primary Still	Oil In	432	215
	Oil Out	430	....
	Top	207	....
Main Still	Oil In	430	18
	Oil Out	407	....
	Top	183	....
De-ethaniser	Feed	169	500
	Top	105	....
	Bottom	245	....
Debutaniser	Feed	136	120
	Top	116	....
	Bottom	287	....
Depropaniser	Feed	124	225
	Top	119	....
	Bottom	202	....
Depentaniser	Feed	77	84
	Top	136	....
	Bottom	141	....
Splitter	Feed	134	28
(Splitting Iso-pentane)	Top	134	....
	Bottom	198	....
Cooling Tower	Water on Collector Deck	64	....
	Water from Process Circuit	76	....
	From Basin or to Turb. Steam Cond.	73	....
	From Turbine Steam Condenser	84	....
Steam	H. P. System	....	380
	Int. P. System	....	90
	Low P. System	....	28

by the equipment mounted above. This additional length at one end of the bottom of the main structure is used to support the main oil heat exchangers.

The condenser and accumulator decks of the structure are open while the ground floor is closed in with corrugated transite walls. All of the steel columns and girders up as high as the second floor are fireproofed with reinforced concrete.

Control instruments for the entire distillation and fractionation processes are concentrated in a room in the center of the ground floor of the structure.

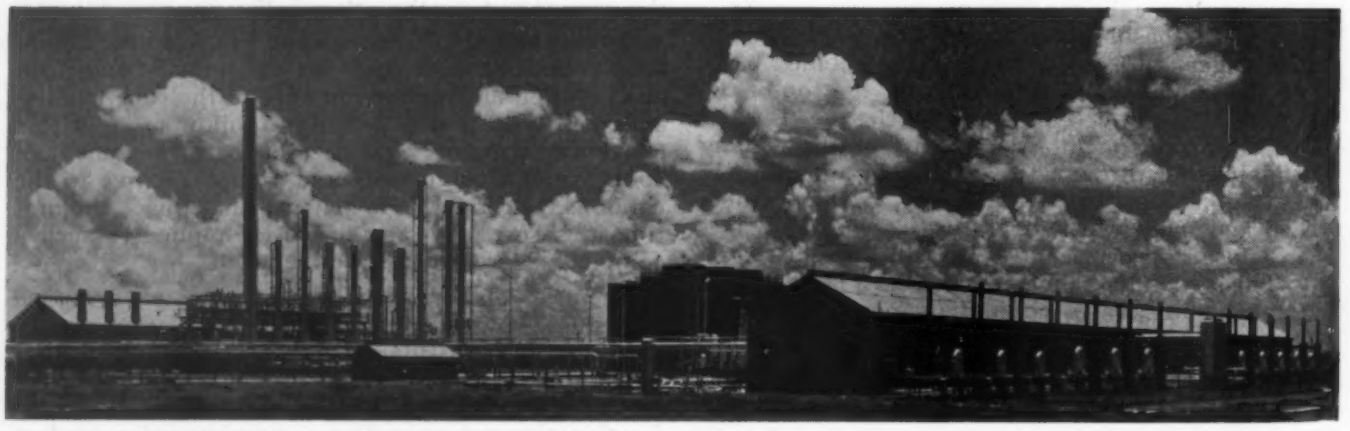
Fire Protection. In addition to the usual hand and cart CO_2 fire extinguishers, the plant is equipped with a dual water fire fighting system.

One of these systems is designed to deliver ordinary hoze nozzle service at the rate of 1500 gallons per minute at a maximum nozzle pressure of 150 psig. Two separate pumping units are provided for this service, one operated by an electric motor and the other by a gasoline engine. In this way nozzle service could be maintained even though electric power were completely out and all natural gas supply cut off from the plant.

The other system is designed to deliver 80 gallons per minute of water at a pressure of 700 psig. for fog nozzle service. This system can be put into operation through any one of five remote control stations at widely separated points around the plant.

Herewith is a tabulation of principal operating conditions in the plant during a test run around the middle of January, 1948.

The plant was engineered, designed and constructed by The Stearns-Roger Manufacturing Company of Denver in collaboration with the Manufacturing Department of Stanolind Oil and Gas Company.



General View of Plant. Left to Right: Boiler and Generator House, Distillation Structure, Cooling Tower, Compressor House.

* * *

Modern Process Methods to Improve L P G Recovery

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IN CONTRAST to the early absorption-type gasoline plants which had but small incentive for extractions greater than required to produce grade gasoline, the present-day plant has a ready market for all propane-butanes at prices attractive to high propane extraction. The knowledge gained during the previous years about the variables influencing absorption now allows wise selection of absorption conditions for economical high propane extraction.

For high propane absorption, a significant saving in lean oil stripping and cooling may be realized by using a flash-flood system as described. In this system only the oil required for pentanes absorption is totally denuded by steam stripping; the balance is flashed at low pressures for partial denudation and introduced at a mid-point in the absorber.

High propane absorption is accompanied by high absorption of ethane, methane, and sometimes hydrogen sulfide. Reduction of these light components by simple venting as applied in the earlier plants is inadequate for economical recovery of the propane and heavier hydrocarbons. Bubble-plate methods improve separation efficiencies and should supplement or supplant venting.

The energy requirements for pumping, stripping, and cooling of absorption oil necessary for high propane extraction and recovery indicates that future plants may make this separation more economically by low temperature liquefaction instead of absorption methods. This would also allow denitrogenation of the residue gas, affording additional advantages by increasing the Btu value of the residue and saving costs of transmitting nitrogen with the residue in pipe lines.

Presentation of this study was made at the March meeting of the Natural Gasoline Association of America at Fort Worth.

THE requisite of the early absorption-oil-type gasoline plants was to extract the pentanes-plus fraction with enough butanes to make a 26-pound product. During the summer months even these butanes were often fractionated off to the plant flare or other distress disposal when the only salable product was a 12-pound butane-free gasoline.

Although there was then but small incentive to study methods for increasing the propane-butanes extraction, the ingenuity of the gasoline plant designer was taxed to provide a minimum cost, low operating expense plant that would yield a fair payout under the natural gasoline prices that existed in the early 1930's. Phase equilibrium constants or, more properly ratios and the absorption factor equation came into widespread use about this time and provided a valuable tool for evaluating the factors influencing absorption.[1, 2, 3] The contributors to this knowledge are many and the data and excellent articles published are too voluminous to mention here.

Under today's market conditions and apparent insatiable demand for liquefied petroleum gas products, the knowledge gained during the earlier but leaner years is being utilized to the utmost to revamp existing plants for increased extraction and to design new plants for almost complete propane extraction or to the maximum allowed by contractual Btu. limitations on the residue gas. The factors influencing absorber performance are well-known but will be briefly reviewed:

1) *Absorber pressure*—The older low pressure plants, operating on rather rich gas compressed from subatmosphere to the range of 40 to 50 psi gage pressure, can now usually pay out additional compressors and higher working pressure absorbers if the gas reserves are adequate for several more years. In the more modern plants the absorber pressure is sometimes higher than the economical optimum for extraction because of pressure maintenance or other residue gas requirements. In design of new plants, the economics of increasing absorber pressure must be compared with other methods of increasing extraction as there is usually an optimum pressure above which the increased extraction may be obtained more economically by other methods.

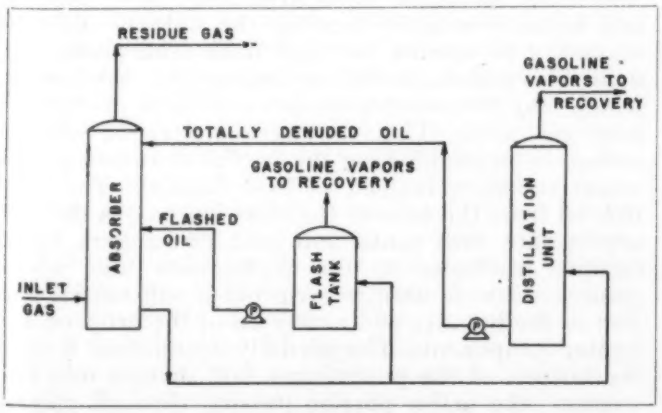

FIGURE 1
Simplified Flow Sheet Flash-Flood System.

2) *Absorption temperature*—The importance of this factor cannot be over-emphasized; it affords a means of increasing extraction in many existing plants by adding more lean oil coils or by supplementing water cooling of the oil with propane refrigeration. An important advantage of the lower absorption temperature is that the heavier constituents are selectively absorbed over the lighter gases. This is an important consideration since the ethane, methane, hydrogen sulfide, and possibly inert gases are detrimental when absorbed in the rich oil and require additional equipment for their elimination. Frequently, increasing extraction of propane-butanes in the absorber fails to net an increase of plant product in the storage tank because the increased extraction of lighter constituents prevents their condensation and recovery.

Additional cooling may be applied directly to the lean oil to the absorber or by using an intercooler to cool the oil a few trays below the top. Lowering the temperature of the lean oil to the top of the absorber results in a lower residue gas temperature, which in turn lowers the moisture content of the residue gas.

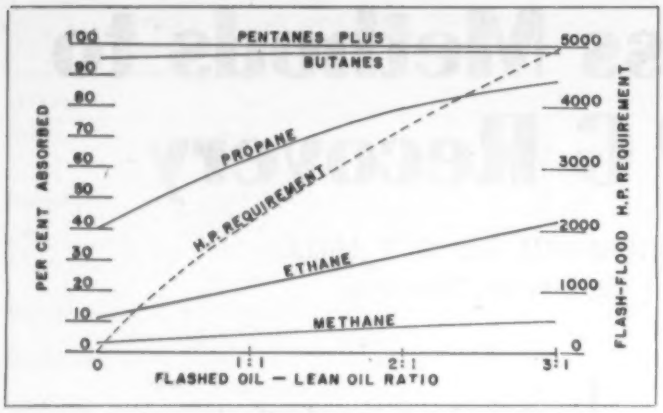

FIGURE 2

Flash-Flood Absorption Characteristics. Lean Oil Rate, 2,300,000 Gallons Per Day; Inlet Gas Rate, 140 Million Standard Cubic Feet Daily; Pressure, 250 Psig.

This may be beneficial in reducing moisture condensation in a pipe line or the load on a gas dehydrator. As approximately a 20° F. reduction in residue gas temperature would halve the moisture content, the investment and operating expense of a gas dehydrator would be substantially lower. When it is not beneficial to lower the residue gas temperature, the intercooler method is more efficient than refrigerating the lean oil as it directs most of the cooling to lowering the effective absorption temperature with less loss of refrigeration to the residue gas. Where there is considerable methane absorption, most of the absorption oil temperature rise occurs near the top of the absorber. The intercooler then removes this heat or absorption and the effective absorption temperature of the oil is benefited by almost the entire amount of refrigeration.

3) *Lean oil molecular weight*—According to the absorption equation, a decrease in lean oil molecular weight results in a proportionate increase in the absorption coefficient but similarly results in a lowering of the stripping coefficient. Between these two factors and other considerations such as oil entrainment, oil retrograde losses, and cost of available absorption oil, the optimum molecular weight oil must be determined. In order to maintain optimum molecular weight, an efficient oil reclaimer is a "must" in present day plants.

4) *Absorber trays*—Like lean oil cooling, increasing the number of absorber trays provides a method of selectively increasing the extraction of desirable constituents over the lighter gases. Many of the older plants are now finding it profitable to increase the number of trays for extraction of more propane and butanes. It is important to provide the optimum number of trays because trays once installed do their work with negligible operating expense.

5) *Lean oil rate*—Most of the other factors mentioned allow a significant but limited increase in extraction while oil rate provides a means of increasing extraction limited only by economics. Every gallon of lean oil circulated requires horsepower for pumping, steam for stripping, and equipment for heating and cooling. It is a costly item and every advantage should be taken to reduce it if economical high propane extraction is to be attained. The characteristics of absorption of any component, as shown by the absorption factor curves, display diminishing returns of increase in extraction for increase in oil rate as the extraction percentage becomes higher. For this reason, the increment of oil rate required to increase propane extraction, say from 60 to 90 percent, results in a greater than proportionate increase in ethane and methane extraction. The absorption coefficient for hydrogen sulfide has been found to fall between the values for propane and ethane. Hence, in areas where the absorber inlet gas is sour, the deep extraction of propane results in high absorption of undesirable hydrogen sulfide. For successful retention of the heavy constituents, the reduction of ethane and lighter with hydrogen sulfide—if present—must be accomplished.

Multiple Absorption Methods

One of the major investment and operating costs of a gasoline plant is that required for equipment to properly strip and cool the lean oil. For absorption of all the pentanes-plus fraction, it is of course necessary that the lean oil be almost completely stripped of pentanes. However, when a gasoline plant is designed for very high propane extraction, the increment of oil required for extraction of propane and butanes is significantly larger than that required for absorption of the pentanes. The principle of multiple absorption is to provide enough completely stripped lean oil to the top of the absorber for absorption of the pentanes and apply at a point lower down in the absorber a larger volume of partially denuded oil for the desired absorption of propane-butanes.[4, 5, 6, 7] A simplified flow sheet is shown in Figure 1, illustrating this principle.

This type of multiple absorption has been termed flash-flood since the oil required for high absorption of the liquefied petroleum gases is circulated at a high rate to the mid-section of the absorber and is partially denuded by cold flashing rather than by steam stripping. Thus the investment for steam stripping and for heating and cooling of the flashed oil is saved. In designing a plant of this type, there are many variables that must be avaluated. The rich oil to the flash tanks may be flashed in any desired number of stages so that the leaner vapors may be segregated for appropriate processing while the major part of the propane and butanes may be flashed off under vacuum and condensed. The number of stages of flashing, the flashing pressures, and the ratio of flashed oil to lean oil are the major variables to be examined.

The curves in Figure 2, show the percentages of hydrocarbon constituents absorbed at varying flashed oil to lean oil ratios at an absorption pressure of 250 psi. gage. Also shown are the compressor horsepower requirements for handling the flashed vapors and for circulating the absorption oil.

In order to attain the desired removal of propane and butanes without heating the flashed oil, it is necessary to operate the final flash tank under vacuum. This makes possible air leakage into the system, which may introduce a serious corrosion problem in sour gas areas. This contingency may be circumvented by accomplishing the partial denudation with steam stripping instead of cold flashing. The total rich oil from the base of the absorber passes through appropriate vent tanks and heat exchangers and a furnace preheater to the preliminary still where enough steam is used to strip out a substantial portion of the butanes and nearly all of the propane and lighter components. The partially denuded oil leaving the bottom of the preliminary still divides into two streams, the major portion passing through cooling coils to the lower section of the absorber, while the other stream—only the quantity required for pen-

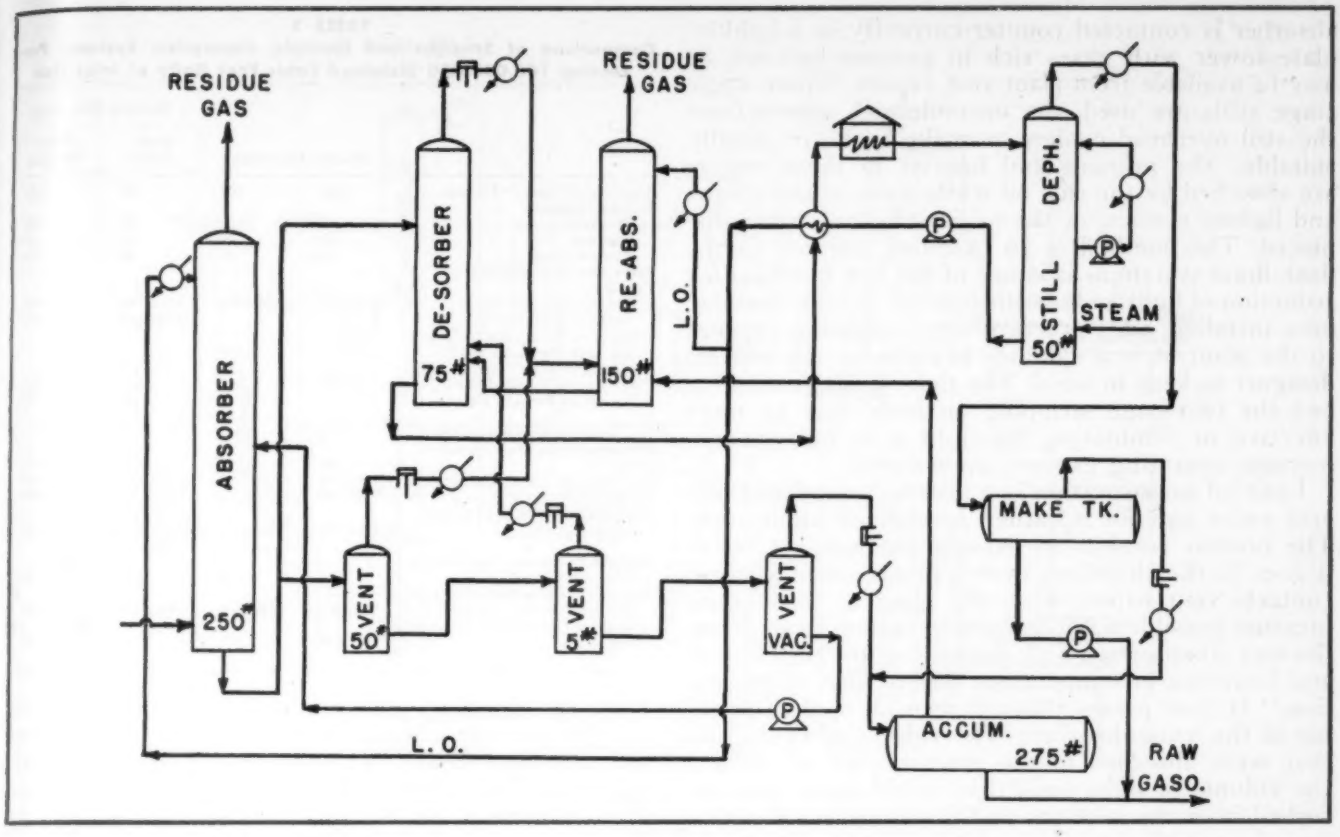

FIGURE 3
Flash-Flood Plant with De-Sorber, No Residue Limitations on Extraction.

tanes-plus absorption—passes to the final still for total denudation.

The investment and operating savings attainable with the partial steam stripping system are not usually as large as with the flash-flood system because the partially stripped oil requires heating and cooling. The savings of this system over straight absorption for equivalent extraction are due to the reduced stripping steam and dephlegmator cooling requirements.

Although the purpose of this paper is to discuss processing for high propane recovery, it should be pointed out while discussing flash-flood multiple absorption that this system offers excellent possibilities for ethane recovery if deired.

Light Gas Removal

The deep absorption of propane is accomplished by high quantities of methane and ethane, the amount depending upon whether high pressure, high oil rate, refrigeration, multiple absorption, or a combination of these methods is used. Although the light-end quantity may be somewhat reduced by advantageous use of refrigeration and absorber trays, even the best selection of variables will ordinarily result in light-end absorption too great for the simple venting methods utilizing pressure reduction and heating of the rich oil, as they would result in high recycle, recompression, and reabsorption costs. Flash drums afford only a single-plate or equilibrium-flash separation, which may be considerably improved by using bubble-plate separation methods to supplement or supplant the "single-plate" flashings. This is analogous to using a bubble-plate fractionator to stabilize raw product instead of a weathering process. The means that the modern designer has devised for more economical separation and recovery of propane-butanes in the light gases are, basically, the use of bubble-plate separations instead of single flashing and include:

1) Rich oil deethanization[8]
2) Two-stage rich oil stripping at high and low pressure[9]
3) Rich oil desorption[10, 11, 12, 13]
4) Lean oil presaturation

Rich oil deethanization is fractionation of the rich oil from the absorbers to eliminate the ethane and lighter. The rich oil from the absorber enters the lower section of a two-section bubble-plate column. The bottom is the stripping section containing a reboiler; the upper section recovers propane and heavier stripped out in the lower section by lean oil absorption. This process was recently described by A. M. Whistler,[8] whose studies indicated this process to be limited to relatively rich gases and for moderate propane recoveries. Recoveries of much higher than 70 percent of the propane in the rich oil are not economical because of the high volume of lean oil required over the top section.

The principle of two-stage stripping is widely applicable to rich oils containing large quantities of light ends as it affords a means of eliminating or greatly reducing costly recompression. The first still should operate at a pressure higher than the reabsorber so that the noncondensable vapors from the overhead cooler may flow to the reabsorber without recompression. Enough steam should be used in the first still so that the overhead vapors from the final still totally condense at the operating temperature and pressure. Although the optimum operating pressures for the two stills depends on the absorber pressure, richness of the inlet gas, and other factors, the order of 150 psi. gage in the first still and 50 psig. in the second illustrates a typical case.

In rich oil desorption the cold rich oil from the

absorber is contacted counter-currently in a bubble-plate tower with gases rich in propane-butanes, as may be available from plant vent vapors. Where single stage stills are used, the uncondensed vapors from the still overhead coolers or make tank are usually suitable. The propane and heavier in these vapors are absorbed by the rich oil while some of the ethane and lighter content of the oil is stripped out or displaced. This method is an excellent adjunct to the flash-flood system as it is one of the few methods for reduction of light ends in the cold oil. As this method, once installed, adds practically no operating expense to the plant, it is a valuable process for the process designer to keep in mind. The rich oil deethanization and the two-stage stripping methods may be more effective in eliminating the light ends but likewise increase operating expense appreciably.

Lean oil presaturation is a process introduced several years ago for a rather specialized application. The process consists in passing the lean oil before it goes to the absorbers over a preabsorber where it contacts vent vapors from the plant at the highest pressure possible without recompression. In so doing, the lean oil absorbs a high quantity of the vent vapors and increases in temperature due to heat of absorption.[14] It then passes through lean oil coolers to the top of the main absorbers. The volume of vent gases that were absorbed in the presaturated oil reduces the volume of light ends that would have been absorbed from the inlet gas and thereby these constituents are, in effect, transferred from the vent vapors to the plant residue. This is accomplished by liquid pumping instead of vapor compression. The removal of much of the heat of absorption by presaturation is equivalent to using an intercooler, thereby reducing effective absorption temperature and increasing the propane absorption. This process is particularly applicable where the main absorber operates at a high pressure to deliver residue gas and where the extraction in the main absorber is limited by Btu. requirements on the residue gas. In effect, the residue gas is enriched by the heavier constituents in the plant low pressure vent vapors.

Selection of Process Flow

There are many factors and variables that must be considered in choosing the process flow for a new extraction plant or revision of an old one, usually requiring that the process designer calculate several alternate cases for selection of the one most desirable. This will usually be the most economical case, but in some instances such factors as flexibility for possible future needs may wield a strong influence.

The major factors that mold the process flow are listed below, omitting the strictly economic factors such as prices received and paid. These economic factors, although affecting the payout and the economic depth of extraction, do not affect the choice of one flow system over another for the same product recovery.

1) Inlet gas volume, pressure, and composition.
2) Residue gas requirements as to volumes, Btu. limitations, temperature, and pressure.
3) Product outlets.
4) Existing equipment (when the project is a plant extension or revamp).
5) Equipment availability.
6) Flow schedule or load factor.
7) Flexibility as to present or possible future operating needs.

A study of the first two items will usually determine the absorber pressure and depth of extraction for the process design. Then several alternate flow schemes may require calculations and economic analysis before selection. If there are no restrictions other than economics on the depth of extraction, it may be necessary to includes cases for several depths of extraction and recovery. A careful study of the factors and flow possibilities may narrow the selection by critical elimination. Likewise, judicious choice of operating conditions will lighten the burden of cases that require calculation. Items four and five should be kept constantly in mind, for they are unpredictable and may upset the labors of many painstaking calculations.

TABLE 1
Comparison of Straight and Multiple Absorption Systems Processing 140,000,000 Standard Cubic Feet Daily of Inlet Gas

	Straight Absorption		Multiple Absorption	
			Flash-Flood	Partial Stripping
Absorption Pressure—Psi gage	250	600	250	250
Percent Absorbed:				
Propane	80	80	80	80
Butanes	99+	99+	98	99+
Pentanes-plus	99+	99+	99+	99+
Absorption Oil Circulation—Gal./Day:				
Totally Denuded Oil	6,350,000	3,000,000	3,500,000	3,150,000
Partially Denuded Oil	.....	.....	4,600,000	3,500,000
Compressor Horsepower for Boosting Inlet Gas above 250 Psi	.....	7,400	.....	.....
Compressor h.p. for Vapor Handling	.....	.....	2,800	.....
Pump h.p. for Totally Denuded Oil	1,200	980	660	600
Pump h.p. for Partially Denuded Oil	.....	.....	900	660
Total Variable Horsepower	1,200	8,380	4,360	1,260
Stripping Steam Requirements—Lb./Hr.	92,500	43,800	51,000	61,000
Oil Cooling Requirements—Btu/Hr.	75,700,000	35,770,000	41,730,000	79,290,000
Oil Heating Requirements—Btu/Hr.	177,000,000	104,500,000	115,800,000	183,500,000
Dephlegmator Cooling Requirements—Btu/Hr.	111,480,000	73,300,000	78,960,000	82,940,000
Relative Investment for Variable Process Equipment	1.47	1.06	1.00	1.40
Annual Operating Expense (Less Gas Royalty and Indirect Exp.)	$383,000	$496,000	$420,000	$363,000
Relative Operating Expense to Flash-Flood	0.91	1.18	1.00	0.87

As an illustration of process choice, assume a case with these premises: 1) An inlet gas volume of 140,000,000 standard cubic feet daily of moderately rich gas with ample reserves for a long plant life; 2) Low pressure field trap conditions such that the gas enters the plant at atmospheric pressure and requires compressors; 3) No residue gas limitations as to pressure or extraction depth; 4) Products to be sold as grade natural gasoline and LPG. This example illustrates a case where the cost of compressing the inlet gas to the absorption pressure must be borne by the gasoline and LPG products recovered without aid from increased value of the residue gas due to pressure. An appropriate flow system is presented in Figure 3.

Some of the economics showing the advantages of the flash-flood system, as given in Figure 3, are compiled in Table 1. For the purposes of this comparison the investments and operating costs for equipment to compress the gas to the 250 psi. gage level are not included since these costs are common to all the cases presented. The tabulated figures show that the flash-flood system of 250 psi. gage absorption pressure accomplishes 80 percent propane extraction at an investment saving over straight absorption at the same pressure, with but $37,000 greater annual operating expense which would require the average lifetime to payout the difference in investment. The same extraction may be accomplished at 600 psi. gage by straight absorption at 6 percent higher investment and 18 percent higher operating expense than for the flash-flood system at 250 psi. gage.

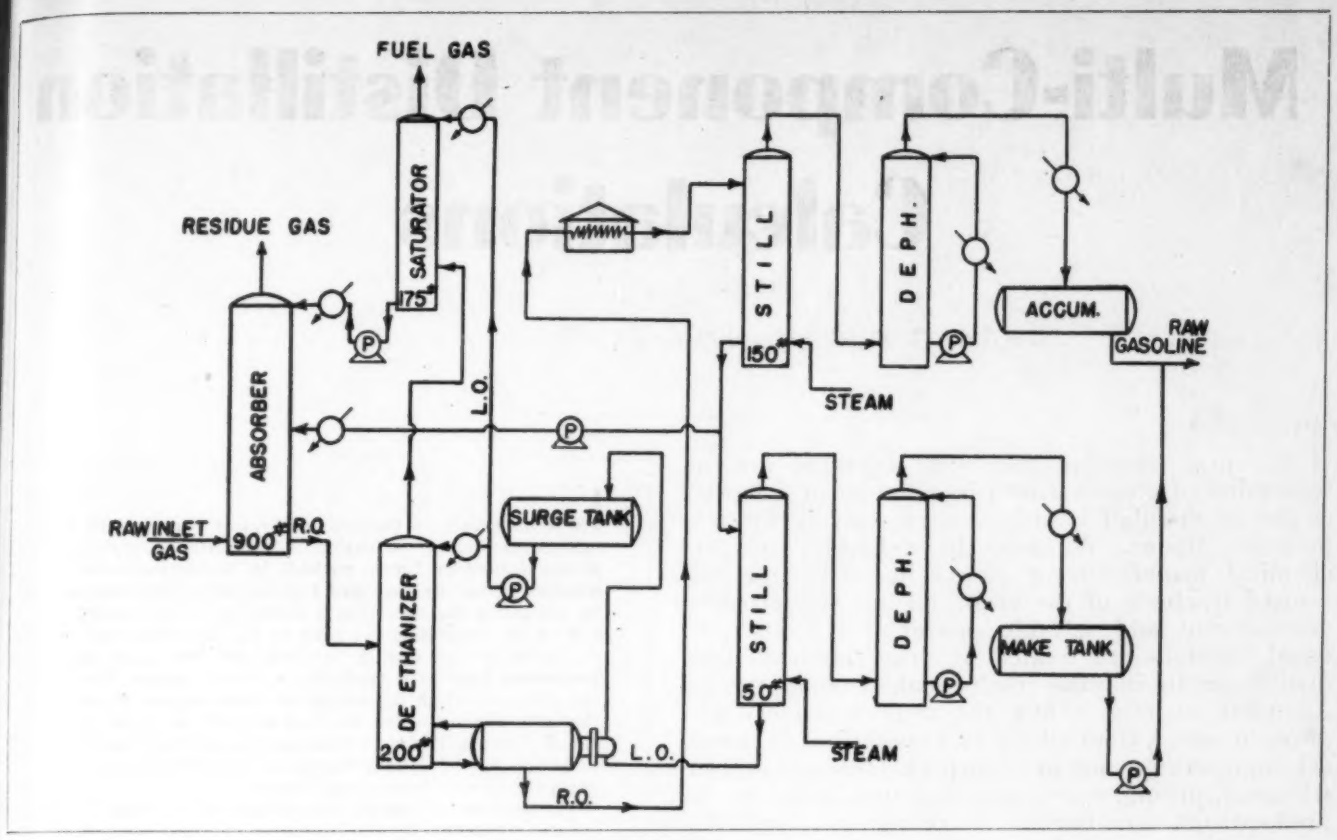

FIGURE 4
Flow Diagram High Pressure Plant, Residue Limitations on Extraction.

Suppose, though, that the inlet gas were available at a high pressure, such as 1000 pounds gage, and that a residue gas contract limited extraction to 50 percent of the propane and all of the butanes-plus. The lean oil requirement for this extraction would be lowered to the extent that flash-flood would not be advantageous. A flow scheme that would be indicated is shown in Figure 4.

The many factors affecting the selection of the most economical plant flow and then selection of optimum operating variables present a complex task to the modern process designer. General rules, formulated for process selection by defined limitations of such factors as inlet gas richness and pressure, become clouded and unsafe by the combination of the many local factors that must be considered for each individual case.

Low Temperature Liquefaction

As a closing thought it is predicted that the present trend toward deep extraction of propane and eventually ethane will alter future plant processes from the absorption-oil type to low temperature liquefaction.[15] The original purpose of the natural gasoline plant was to remove the "wet" or heavy gasoline ends but the present day trend is to make the separation not at the bottom but at the top, removing the light gases from the liquefiable hydrocarbons. This objective, it seems, should be accomplished more efficiently by straight low temperature fractionation of the inlet gas. The high energy requirements for pumping, heating, stripping and cooling of the absorption oil indicate promise for low temperature liquefaction separation methods, which are thermodynamically more direct. The introduction of any foreign medium, such as absorption oil or charcoal, increases the energy requirement for separation.

In some areas where extraction is limited by Btu. limitations on the residue and where there is appreciable nitrogen in the gas, low temperature liquefaction offers the advantage of nitrogen separation. Removal of nitrogen would allow deeper extraction of propane and ethane, if desired and would save the pipeline cost of transmitting the inert nitrogen. Not only could horsepower be saved for transmitting the inert material, but existing pipeline and compressor installations could be made to handle more therms.

Although the process of low temperature liquefaction has not been completely developed, representatives of both the natural gasoline and gas transmission industries must recognize that cooperative approach to its use will be required if denitrogenation and hydrocarbon recovery are to be attained most economically.

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Multi-Component Distillation Calculations

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IN THIS article, a thorough analysis is made of all the factors involved in the design of a column for multicomponent distillation. Simple methods for the determination of minimum reflux are described together with simple methods for evaluating the approximate number of trays required. A method for calculating the ratio of the key components on the optimum feed tray is discussed, and the equations of Underwood have been applied to a simple tabular form of calculation in which the number of trays required to obtain this ratio of the keys on the feed tray can be easily determined. The application of Underwood's equations have been simplified by the introduction of a function G, described, which can be read from a simple graph.

The problem of heat balances around all the trays in the column to determine the variation of vapor and liquid quantities is also discussed, and a new method for adapting the Ponchon method to multicomponent distillation is described. This greatly simplifies the exact design of multicomponent distillation column including heat balances on all the trays.

THE first refineries that were operated for the separation of crude oil into its component fractions employed distillation processes exclusively for this purpose. Modern refineries have become in part chemical manufacturing plants in which the unwanted fractions of the crude oil are converted to more useful and saleable products, and also the usual fractions are treated by other methods than distillation to improve their quality or usefulness for other purposes. Thus, the modern refinery employs to some extent all the unit operations of chemical engineering, such as absorption, adsorption, centrifuging, drying, extraction, filtration, etc., but the predominant operation in all refineries is still distillation.

The earliest design of distillation equipment was by "rule of thumb." However, with the introduction of rigid and continually improved specifications for petroleum products, it was necessary to develop a sounder basis for design. The first requirement in any design is a knowledge of physical and physical chemical properties of the components and their mixtures, and up to the present time, much more work has been done on the pure hydrocarbons, and their mixtures with each other and with other compounds, than on any other groups of compounds. Thus, while experimental techniques are still being improved in this field, the basic physical chemical data have been determined with a good degree of accuracy.

The method of application of these data to the design of equipment handling multicomponent mixtures has also been the subject of considerable work in recent years, stimulated by the close competition in the field of petroleum process equipment design and construction. Most of this work has been directed toward predicting from the observed data on the pure components, of which there are a limited number, the properties and performance of multicomponent mixtures, of which there are unlimited numbers. There is very little that cannot be calculated for the properties of hydrocarbon mixtures from the data on the pure components, although the reliability of the final answer varies, and in many of these properties greater accuracy is both possible and desirable.

Mixtures Obey Raoult's Law

The basic data for the design of distillation equipment is fairly well established in the ordinary range of temperatures and pressures used in this equipment, although it is not yet possible to accurately predict the distillation equilibria of a multicomponent mixture at temperatures and pressures approaching the critical conditions of the mixture.

The calculation of the vapor and liquid concentration in equilibrium in a hydrocarbon mixture is relatively simple because these mixtures are very nearly ideal and obey Raoult's law which states that the partial pressure of a component from a mixture is proportional to the concentration of the component in the mixture, and the proportionality constant is independent of the concentration. Under these conditions, the proportionality constant is equal to the vapor pressure of the pure component. When allowance is made for the deviations of the gas and liquid properties from the perfect gas laws by the use of fugacities, the proportionality constant is called the K-value. A knowledge of these K-values for all the various hydrocarbons at different temperatures and pressures is the primary requisite for the design of distillation equipment for petroleum hydrocarbon mixtures. These K-values have been reproduced for all of the normal paraffin hydrocarbons at different temperatures and pressures on a single chart.[14, 15] The ratio of the K-values of any two components is called their relative volatility and, while K-values vary considerably with temperature at a given presture, the relative volatility varies very little with temperature. It can be shown from the Clausius Clapeyron equation that the variation of the relative volatility of two components with temperature is a function of the difference between their molal latent heats. Thus, it should be noted that only when the two components have identical molal latent heats, is their relative volatility constant, and since no two hydrocarbons satisfy this condition, truly constant relative volatility is never realized, even in ideal solutions. However, in most cases, the variation of the relative volatility over the temperature range of an

ordinary distilling column is sufficiently small so that the use of a constant value does not introduce any large error in the calculation.

In the design of a distilling column, it is necessary that the reflux ratio be greater than a certain minimum value which would require an infinite number of trays to give a specific separation and in practice, an economical design would be at a reflux ratio 1.5 to 2.0 times the minimum value. When the reflux ratio is chosen, it is then necessary to determine the number of trays required to give the separation desired in the column. For preliminary and economic studies, it is possible to estimate the total trays required and a final design involves a rigorous calculation of the number of trays required in the column. Thus, a complete process design of a distilling column consists of the following steps:

1) Determination of minimum reflux ratio.
2) Estimation of the number of trays required at a reflux ratio greater than the minimum.
3) Calculation of the exact number of theoretical trays required in the column.

Each of these steps will be considered separately in arriving at the final column design.

MINIMUM REFLUX RATIO

The minimum reflux ratio for a binary system can be readily calculated or determined graphically from the McCabe-Thiele diagram[12] since it is that reflux ratio which gives an intersection of the operating lines on the equilibrium curve of the two components. In multicomponent distillation, the situation is not so simple because the equilibrium curve varies with each feed composition and with each reflux ratio so a direct solution is not possible by this method.

The tray calculations for a binary system at a given reflux ratio reach a constant composition of vapor and liquid (given by the intersection of the operating line with the equilibrium curve) and a similar effect is observed with multicomponent calculations although its graphical significance is not apparent. Fenske[5] and Underwood[19] have derived equations for directly calculating the composition of this so-called "pinch" at which point the compositions of the vapor and liquid do not change with additional tray calculations. At minimum reflux in a binary distillation, the pinch compositions for the stripping section and the fractionating section of the column are identical, and on this basis methods for relating these pinch compositions at minimum reflux in multicomponent distillation have been developed.

When separating several components in a distillation column, it is necessary to specify the maximum concentration of one of the components that is desired in the overhead product and the maximum concentration of one of the components, usually the next more volatile component, in the bottoms product. These components are referred to as the key components, the first being the heavy key and the second, the light key. If a component is present which has a volatility intermediate between the key components, it is called a "split key" and the distillation in this case is then called a "sloppy" separation to distinguish it from the first case which is considered a sharp separation, although the degree depends on the actual concentrations of these key components in the overhead and bottoms products.

The fractionating pinch at minimum reflux contains only the components present in the overhead and, consequently, no components heavier than the heavy key and conversely the stripping pinch contains only the components in the bottom and no components lighter than the light key. The first exact definition of minimum reflux ratio in a multicomponent system was given by Jenny.[8] When a trace of the components lighter than the keys is introduced into the stripping section pinch and a trace of the components heavier than the keys is introduced into the fractionating pinch, at the minimum reflux the tray calculations between the pinches will be found to match the concentrations of all components at the feed tray. This is a rigorous but very tedious procedure and several attempts have been made to arrive at an equally reliable answer with much less work.

Tray Calculations Eliminated

The earliest methods for estimating minimum reflux ratios was by determining that reflux ratio at which the ratios of the key components in the two pinches were equal. Colburn[4] showed that this was not true and developed an empirical method for determining the relation between the ratio of the key components in the two pinches at minimum reflux. This method involved a trial and error solution in which the ratios of the keys were calculated in both pinches at different reflux ratios until the empirical relationship between them had been satisfied. It was much simpler than the rigorous Jenny method since it eliminated the necessity of tray calculations and gave a value that agreed with the true one with an average deviation of 1 percent over the range of concentrations studied.

Mayfield and May[10] developed a simpler method for calculating the minimum reflux ratio for an essentially complete separation of a multicomponent mixture in which one of the products consists only of one component and a negligible quantity of the other key. The method thus applies only to distillations in which no components are present lighter than the light key or in which no components are present heavier than the heavy key.

An empirical equation[17] for direct calculation of the minimum reflux ratio for a multicomponent mixture has also been developed which is based on the concept of a pseudo minimum reflux ratio. This pseudo minimum reflux ratio is that required to separate the key components when all components heavier than the keys have zero volatility and all components lighter than the keys have infinite volatility. This value is readily determined from the binary equilibrium curve of the key components by constructing a feed line on the McCabe-Thiele diagram with a negative slope equal to the pseudo ratio of liquid to vapor in the feed which is given by the following relationship:

$$m = \frac{X_L - \Sigma X_D}{X_V - \Sigma X_A} = \frac{M_L - \Sigma M_D}{M_V - \Sigma M_A} \qquad (1)$$

where m = pseudo ratio of liquid to vapor in feed
X_L = mole fraction of feed as liquid
X_V = mole fraction of feed as vapor
ΣX_D = total mole fractions of components heavier than heavy key in feed
ΣX_A = total mole fractions of components lighter than light key in feed
M_L = moles of liquid feed
M_V = moles of vapor feed
ΣM_D = total moles of components heavier than heavy key in feed
ΣM_A = total moles of components lighter than light key in feed

The intersection of the feed line with the equilibrium curve is most easily determined by the graphical

method. However, when the binary equilibrium curve is not readily available but the relative volatility is known, an equation can be derived expressing the value of the intersection in terms of the known quantities. The equation is cumbersome but it gives a more precise value than the graphical solution unless a large scale is used for the equilibrium curve. However, where repeated calculations using the same two key components are made, their equilibrium curve can be kept available for such constructions and the graphical treatment is well within the accuracy required in any industrial calculation.

The pseudo minimum reflux ratio is then calculated by the following equation:

$$R'_M = \frac{x_F}{(\alpha_B - 1)x_i} - \frac{(1 - x_F)\alpha_B}{(1 - x_i)(\alpha_B - 1)} \qquad (2)$$

where R'_M = pseudo minimum reflux ratio
x_F = concentration of light key in the overhead product expressed as a fraction of the total amount of the keys
x_i = abscissa of the intersection of the feed line with the equilibrium curve
α_B = relative volatility of light key with respect to heavy key corresponding to the concentration of x_i on the binary equilibrium curve.

When the pseudo minimum reflux ratio has been determined, it is possible to correct it for the actual volatilities of the components heavier than the keys and lighter than the keys by the following empirical equation:

$$R_M = \frac{1}{X_B + \Sigma X_A}\left[X_B R'_M + (X_C + \Sigma X_D) \sum \frac{X_D}{\frac{\alpha_B}{\alpha_D} - 1} + \sum \frac{X_A}{\alpha_A}\left(1 + \frac{\alpha_B}{\alpha_A}\right)\right] \qquad (3)$$

where X_A = concentration of each component lighter than the keys in the feed
α_A = relative volatility of each component lighter than the keys with respect to the heavy key
X_D = concentration of each component heavier than the keys in the feed
α_D = relative volatility of each component heavier than the keys with respect to the heavy key

This equation can be expressed in terms of the moles of feed and the K-values as follows:

$$R_M = \frac{1}{M_B + \Sigma M_A}\left[M_B R'_M + \frac{M_C + \Sigma M_D}{M_F} \sum \frac{M_D}{\frac{K_B}{K_D} - 1} + \sum M_A \frac{1 + \frac{K_B}{K_A}}{\frac{K_A}{K_C}}\right] \qquad (4)$$

where M_A = moles of each component lighter than keys in feed
M_B = moles of light key in feed
M_C = moles of heavy key in feed
M_D = moles of each component heavier than keys in feed
M_F = total moles of feed
K_A, K_B, K_C, K_D are the K-values of each of the above components at the temperature of the feed tray for the pseudo minimum reflux ratio based on the binary system of the keys.

These equations can be applied directly to the determination of minimum reflux ratio in a multicomponent system without the usual trial and error solution, and as such they are simpler than other published methods. The equation has been found to agree with the true value with an accuracy of about 1 percent for the cases usually encountered in practice and in extreme cases where the key component comprise only 10 percent of the total feed, the agreement is within 10 percent of the true minimum reflux.

Recently Bailey and Coates[1] pointed out that larger errors might be obtained with mixtures where the relative volatility varies considerably in the column, as in cases where the feed contains a large quantity of components either lighter than the keys or heavier than the keys. They developed a more reliable method for determining the pseudo minimum reflux ratio in these cases and also found that by calculating the first correction term at the temperature of the stripping section pinch, and the second correction term at the temperature of the fractionating section pinch, this equation can be applied to these cases with an accuracy of 1 percent.

Underwood[21, 22] has recently derived a rigorous method for calculating this minimum reflux ratio and while it involves the solution of a rather complex equation it is possible, by the use of a relatively simple trial and error technique described later, to obtain the desired root of the equation from which the minimum reflux ratio can be readily calculated. Thus, this method gives an exact value for the minimum reflux ratio with only a small amount of additional work.

Estimation of the Number of Trays Required at a Given Reflux Ratio

When the minimum reflux ratio has been established, it is possible to choose an operating reflux somewhat larger and determine the number of theoretical trays required. However, it is frequently desirable to know the approximate number of trays required for preliminary process and economic studies. This can be obtained from general correlations[3,7] based on the minimum reflux ratio and the minimum number of trays, i.e., those required at total reflux.

The trays required at total reflux are readily calculated from the equation of Fenske[5] and of Underwood[19] which can be readily applied to multicomponent mixtures in the form:

$$n + 1 = \frac{\log \frac{x_B x'_C}{x'_B x_C}}{\log \alpha} = \frac{\log \frac{m_B m'_C}{m'_B m_C}}{\log \alpha} \qquad (5)$$

where x_B = concentration of light key in overhead product
x_C = concentration of heavy key in overhead product
x'_B = concentration of light key in bottoms product
x'_C = concentration of heavy key in bottoms product
m_B = moles of light key in overhead product
m_C = moles of heavy key in overhead product
m'_B = moles of light key in bottoms product
m'_C = moles of heavy key in bottoms product
n = total number of trays in column operated with reboiler and total condenser.

For the case of a partial condenser with vapor overhead product, the value of the right hand side of the equation is $n + 2$.

As a very approximate rule, the number of trays required at 1.5 times the minimum reflux ratio is about twice the minimum number of trays and at twice the minimum reflux the trays required will be about one and a half times the minimum number. The first general correlation between the number of trays and the reflux ratio was developed by Brown and Martin.[8] Gilliland[7] subsequently proposed a correlation which was simpler to apply and gave results of the same degree of accuracy. These correlations

give results that are usually within 10 percent of the correct answer.

The previous correlations predict only the total number of trays required and do not give any clue to the location of the feed tray. The concept of the pseudo minimum reflux ratio has been extended to the pseudo reflux ratio[13] which is defined as that reflux which gives the same total number of trays on the binary equilibrium diagram of the key components as required in the multicomponent separation. It has been found that the pseudo minimum reflux ratio can be calculated from the following empirical relationship:

$$R' + x_1 = (R + x_1)\,\frac{R'_M + x_1}{R_M + x_1} \qquad (6)$$

where
R' = pseudo reflux ratio
R = actual reflux ratio

By using the pseudo reflux ratio to carry out the graphical tray calculations on the binary equilibrium curve of the key components by the McCabe-Thiele method,[12] it is possible to locate the feed tray and also determine the total number of trays in the column with an accuracy better than 10 percent and in most cases within 5 percent of the correct value.

Bailey and Coates[2] modified this equation and proposed a method for calculating a modified binary equilibrium curve depending on the composition of the feed. They were able to obtain an agreement with the rigorous tray-to-tray calculations of within 3.5 percent in all cases studied.

CALCULATION OF EXACT NUMBER OF TRAYS REQUIRED IN A MULTICOMPONENT DISTILLATION:

The rigorous method of tray-to-tray calculations as described by various authors involves a trial and error matching of all the components at the feed tray and, as such, it is a difficult, time-consuming process in which a completely exact answer is almost impossible, and the concentrations of the components are generally matched with such a degree of accuracy that a perfect matching of all the components would not appreciably effect the number of trays required. Jenny[8] has proposed a method in which the tray calculations can be started at the feed tray and thus, the trial and error matching of the components is substantially eliminated. The method, of course, requires only a fraction of the time required by the first method, but it is also time-consuming and requires one complete set of tray calculations for an accurate solution, even after the compositions of the feed tray liquid and vapor are established. Other methods of carrying out multicomponent tray calculations have been proposed but these introduce uncertainties due to the approximations involved in shortening the method, and cannot be considered any more reliable than the approximate methods of the previous section although they require more calculations.

Recently Underwood[20, 23] has derived equations by which it is possible to calculate the concentrations of all components on a tray in multicomponent calculations and these equations can be adapted to a convenient form of calculation which is no more difficult or time-consuming than the ordinary dew point and bubble point calculations made on each tray for multicomponent distillation. The method requires a trial and error procedure for matching the components at the feed tray. However, this can be easily done very accurately because the concentrations are determined by a single calculation and not by a series of tray-to-tray calculations as in previous methods. The method completely eliminates the tray to tray calculations because the feed tray compositions are obtained directly from the equation.

A) Location of Optimum Feed Tray:

In multicomponent tray calculations the optimum feed tray location is not as obvious as in the case of a binary system where the optimum feed tray is near the intersection of the operating lines. The feed tray for any multicomponent distillation can be located any place between two limiting positions each of which require an infinite number of trays and the optimum feed tray location is defined as that which requires a minimum number of total trays in the column. This would require an additional number of calculations at different feed tray locations to establish the optimum. The method of Gilliland[6] for choosing this optimum feed tray location does not hold and the ratio of the key components on the optimum feed tray has been found to vary with the reflux ratio.[17] An empirical equation has been developed to calculate the ratio of the key components on the feed tray at the minimum reflux ratio.[17] The ratio of the key components in the optimum feed tray at any reflux ratio has been found to vary linearly with the function $\frac{R - R_M}{R}$ between the limiting values at minimum and total reflux. The optimum feed tray approaches at total reflux, the tray which straddles the feed composition such that the ratio of the key components is equal to the ratio of these components in the feed divided by the square root of their relative volatility. This ratio has been confirmed by precise calculations on binary systems. It is apparent that the optimum feed tray at total reflux is the same for all conditions of the feed.

Also at minimum reflux the ratio of the key components in the feed tray (there is only one feed tray possible at minimum reflux) is given by the empirical equation:

$$r_M = \frac{r'_M}{1 + F_1 \sum \alpha_D (1 - \alpha_D)\, X_D + F_2 \sum \frac{X_A^2}{(\alpha_A - 1)} - F_3 \sum \frac{(\alpha_A - \alpha_B)\, X_A^2}{(\alpha_A - 1)^2}} \qquad (7)$$

Where r'_M = ratio of light to heavy key in feed tray at pseudo minimum reflux ratio

$$r'_M = \frac{x_1}{1 - x_1} \qquad (7a)$$

r_M = ratio of light to heavy key in optimum feed tray at actual minimum reflux ratio

$$F_1 = \frac{(1 - \alpha_B v)\, \alpha_B X_B}{(\alpha_B - 1)\,(X_B + X_C)\,(1 - \Sigma X_A)} \qquad (7b)$$

$$F_2 = \frac{(\alpha_B - 1)^2 l}{(X_B + X_C)\,(1 - \Sigma X_D)} \qquad (7c)$$

$$F_3 = \frac{(\alpha_B - 1)\, v}{\alpha_B\,(X_B + X_C)\,(1 - \Sigma X_D)} \qquad (7d)$$

l = fraction of feed as liquid
v = fraction of feed as vapor

In this equation it should be noted that all the terms in the denominator are dimensionless quanti-

ties and any consistent units for concentration may be used. Thus all the values of X may be substituted by the corresponding values of M and the terms $(1-\Sigma X_A)$ and $(1-\Sigma X_D)$ must then be substituted by $(M_F-\Sigma M_A)$ and $(M_F-\Sigma M_D)$ respectively.

Also at total reflux

$$r_t=\frac{X_B}{X_C\sqrt{\alpha_B}}=\frac{M_B}{M_C\sqrt{\alpha_B}} \qquad (8)$$

and at any intermediate reflux ratio it can be shown that

$$r=r_t+\frac{R_M}{R}(r_M-r_t) \qquad (9)$$

r = ratio of light to heavy key on optimum feed tray at reflux ratio R.

The value of r determined by this method has been found to locate the optimum feed tray within one to two tenths of a tray and this was considered almost within the limits of accuracy of the graphical technique for locating the optimum feed tray at the minimum number of total trays required because in no cases were the total trays at the calculate value of r any greater than at the previously estimated ratio. Thus the curve of total trays against feed tray location is usually sufficiently flat, particularly at reflux ratios 1.5 times the minimum reflux ratio that the calculated value of r gave perfect agreement with the minimum number of trays even when the total trays were interpolated to within tenths of a tray.

B) Tray Calculation for Systems of Constant Relative Volatility

Using the value of r calculated by these equations it is possible to apply the relationships derived by Underwood to determine the number of trays required above and below this feed tray to reach this ratio. Underwood's equations when written algebraically appear very complicated however it must be remembered that an algebraic expression for a dew point or bubble point calculation also appears very complicated whereas the actual arithmetic solution is very simple when carried out in tabular form. Underwood's equations similarly are no more difficult to apply in tabular form than these other calculations and the easiest method for demonstrating the application of the equations is by considering a specific problem. In order to adequately demonstrate the method but without introducing excessive complications, a four component system will be considered with one component lighter than the keys and one component heavier than the keys. Then in the case of additional components lighter and heavier than the keys these are handled in a similar manner but of course with additional calculations. In general, the components in a petroleum distillation such as ethane, propane, butane, pentane, etc., frequently have relative volatilities of the order of 2 with respect to the next less volatile component in the mixture so relative volatilities of 4, 2, 1 and .5 will be taken for the four components of the feed mixture. Components having higher or lower volatilities with respect to the heavy key are actually easier to handle in the calculations, so for purposes of this illustration they will not be considered. Thus the following separation is desired:

	FEED		OVERHEAD PRODUCT		BOTTOMS PRODUCT	
	Moles/Hr.	α	Moles/Hr.	Mole Fraction	Moles/Hr.	Mole Fraction
A	30	4	30.0	.60	....	....
B	20	2	19.5	.39	.5	.01
C	20	1	.5	.01	19.5	.39
D	30	.5	....	.....	30.0	.60
	100	....	50.0	1.00	50.0	1.00

The feed will be taken as total liquid so that the vapor quantities above and below the feed are the same and constant molal downflow will be assumed in the two sections of the distilling column for these calculations. The significance of this will be considered later.

The first step in the solution of this problem is the determination of the minimum reflux ratio. Thus by equation (1)

$$m=\frac{100-30}{-30}=-2.33$$

By using the equilibrium curve for the key components or plotting it if a curve for $\alpha=2$ is not available, according to the relationship,

$$y=\frac{\alpha x}{1+(\alpha-1)x}$$

it is possible to construct a line with a negative slope of m or in other words a positive slope of 2.33 as shown in Figure 1. The intersection of this line with the equilibrium curve is found as

$$x_t=.610$$

and the pseudo minimum reflux ratio is calculated by equation (2) using

$$x_p=\frac{.5}{39.5}=.975$$

$$R'_M=\frac{.975}{(2-1)\ .610}-\frac{(.025)2}{(.390)\ (2-1)}$$

$$=1.600-.128=1.472$$

The minimum reflux ratio for the multicomponent distillation can be calculated from equation (4) expressed in terms of relative volatilities instead of the K-values

$$R_M=\frac{1}{20+30}\left[(20)\ 1.472+\frac{20+30}{.100}\left(\frac{30}{2/.5-1}\right)+\frac{30}{4}(1+2/4)\right]=$$

$$\frac{1}{50}(29.4+5.0+11.2)=.912$$

In this system, the tray-to-tray calculations indicated the minimum reflux ratio as .89, and the error in this case is more than 2 percent which is larger than in most of the other cases studied.[17] It is desired to operate the column at about twice the minimum reflux ratio and 1.78 was chosen.

In order to determine the approximate number of trays required, the graphical tray calculations are carried out as shown on Figure 1 at the pseudo reflux ratio calculated by equation (6).

$$R'=(1.78+.61)\frac{1.47+.61}{.91+.61}-.61=3.27-.61=2.66$$

It can be shown that for a binary system, the optimum feed tray varies linearly on this diagram between that at minimum reflux determined by the

intersection of the feed line and the equilibrium curve and that at total reflux calculated by the equation

$$x_T = \frac{x_f}{x_f + (1 - x_f)\sqrt{\alpha}} \qquad (8a)$$

where

x_T = concentration of light key in optimum feed tray liquid at total reflux

$$x_f = \frac{X_B}{X_B + X_C}$$

Thus, the graphical tray calculations are started at the intersection of the operating line for the stripping section with the line MT shown in Figure 1, and the trays required are found by this method to be 7.1 in the fractionating section and 8.6 in the stripping section for a total of 15.7 trays. Actual tray-to-tray calculations on the multicomponent column gave 7.0 in the fractionating section and 7.3 in the stripping section for a total of 14.3. Thus an error of about 10 percent is observed with this method. This is also larger than in most other cases studied. However, if the correct minimum reflux ratio had been used to calculate the pseudo reflux ratio instead of the calculated minimum, the agreement would have been better.

It is also interesting to note that the very rough rule based on the trays at total reflux holds exactly for this case, although such close agreement is not often encountered. The minimum number of trays calculated by equation (5) is

$$n + 1 = \frac{\log \frac{19.5}{.5} \frac{19.5}{.5}}{\log 2.00} = \frac{3.184}{.301} = 10.58$$

Thus, n = 9.58 and at twice the minimum reflux ratio, the number of trays required will be about 1.5×9.58 or 14.37. This method does not give any clue as to the location of the feed tray.

A final design would require a more rigorous calculation at the optimum feed tray location. The ratio of the key components on this feed tray is determined by first calculating the ratio of the key components on the feed tray at minimum reflux by equations (7a) (7b) (7c) (7d) and substituting in equation (7) written in terms of the moles of each component in the feed.

$$r'_M = \frac{.610}{1 - .610} = 1.565$$

$$F_1 = \frac{(1-0)\,(2)\,(20)}{(2-1)\,(40)\,(70)} = .01428$$

$$F_2 = \frac{(2-1)^2}{(40)\,(70)} = .000357$$

$$F_3 = 0$$

$$r_M = \frac{1.565}{1 + .01428\,(.5)\,(1-.5)\,30 + .000357\,\frac{30^2}{4-1}} = \frac{1.565}{1 + .107 + 1.07} = 1.29$$

The value of r_T is calculated from equation (8) as follows

$$r_T = \frac{20}{20\sqrt{\alpha}} = .707$$

and the value of r at a reflux ratio of 1.78 is calculated

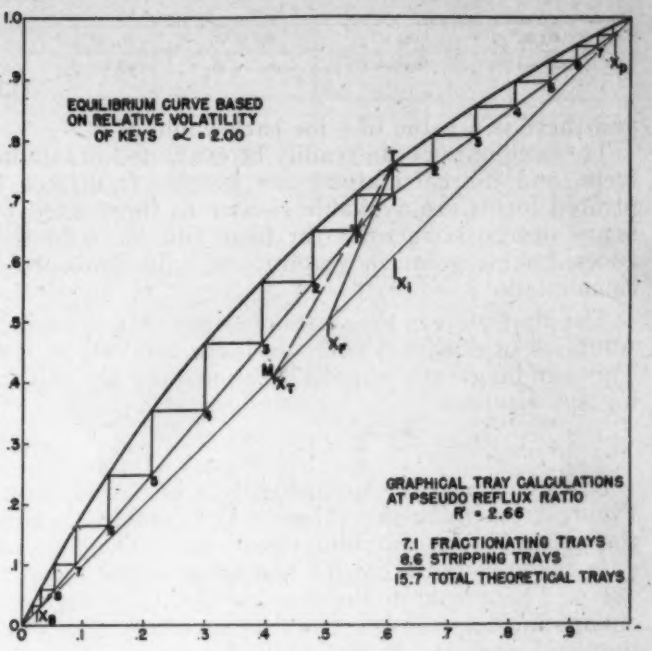
EQUILIBRIUM CURVE BASED ON RELATIVE VOLATILITY OF KEYS ∝ = 2.00
GRAPHICAL TRAY CALCULATIONS AT PSEUDO REFLUX RATIO R' = 2.66
7.1 FRACTIONATING TRAYS
8.6 STRIPPING TRAYS
15.7 TOTAL THEORETICAL TRAYS

FIGURE 1

from equation (9)

$$r = .707 + \frac{.912}{1.78}(1.29 - .707) = .707 + .299 = 1.006$$

It should be noted that, while this value is practically the same as the ratio in the feed, this is entirely coincidental because at other reflux ratios, the value of r varies between .707 and 1.29.

Underwood[23] has derived the equations for the concentration of each component on any fractionating tray in a multicomponent distillation involving k components as follows:

$$x_{An} = \frac{\sum_{i=1}^{i=k} \frac{\lambda_i x^\circ_{Ai}}{\phi_i^n}}{\sum_{i=1}^{i=k} \frac{\lambda_i}{\phi_i^n}} \qquad (10a)$$

and

$$x_{Bn} = \frac{\sum_{i=1}^{i=k} \frac{\lambda_i x^\circ_{Bi}}{\phi_i^n}}{\sum_{i=1}^{i=k} \frac{\lambda_i}{\phi_i^n}} \quad \text{etc.} \qquad (10b)$$

where

ϕ is a solution of the equation

$$\frac{\alpha_A x_{AP}}{\alpha_A - \phi} + \frac{\alpha_B x_{BP}}{\alpha_B - \phi} + \frac{\alpha_C x_{CP}}{\alpha_C - \phi} + \frac{\alpha_D x_{DP}}{\alpha_D - \phi} + \ldots = R + 1 \qquad (11)$$

where the first subscript after the x terms refers to the component and the second subscript refers to the overhead product. This equation will have as many solutions as there are components in the mixture and the subscript i refers to each of these solutions.

Also

$$x^\circ_{Ai} = \frac{x_{AP}\,\phi_i}{R\,(\alpha_A - \phi_i)} \qquad (12A)$$

$$x^\circ_{Bi} = \frac{x_{BP}\,\phi_i}{R\,(\alpha_B - \phi_i)} \qquad (12B)$$

etc.

and

$$\lambda_1 = \frac{1}{\frac{\alpha_A x^\circ_{A1}}{\alpha_A - \phi_1} + \frac{\alpha_B x^\circ_{B1}}{\alpha_B - \phi_1} + \frac{\alpha_C x^\circ_{C1}}{\alpha_C - \phi_1} + \frac{\alpha_D x^\circ_{D1}}{\alpha_D - \phi_1} + \ldots} \quad (13)$$

and there is a value of λ for each value of ϕ.

These equations can readily be evaluated in tabular form, and the calculations are greatly facilitated if printed forms are available similar to those used by many design companies for heat and material balances, bubble point, dew point, and equilibrium flash calculations.

The first step is the determination of the various solutions of equation (11) to obtain the values of ϕ. This can be greatly simplified by making the following substitution

$$\text{Let } G = \frac{\alpha - \phi}{\alpha} = 1 - \frac{\phi}{\alpha} \quad (14)$$

It is then possible to construct a chart similar to Figure 2 in which the value of G is plotted against the value of ϕ at different values of α. The lines all pass through the point 0,1 and have slopes equal to $-1/\alpha$. The break in the line at $G = 0$ is due to a change in the scale because it is necessary that positive and negative values of G be determined with equal accuracy and it is apparent that this could not be done by using the same scale for positive and negative values of G. In spite of the change in the scale, both segments of the lines are straight and convergent. A somewhat more convenient graph may be constructed by plotting log (1—G) vs log ϕ on regular logarithmic graph paper and calibrating the ordinate directly in G as shown in Figure 2(a). If the inverted calibration of the G scale is considered confusing, the lettering may be made on the reverse side of the graph paper and the chart printed through the paper.

Equation (11) may then be written

$$\sum \frac{x_P}{G} = R + 1 \quad (11a)$$

where the summation refers to all the components present in the column and x_P refers to their individual concentrations in the overhead. By a trial and error procedure values of G for the given values of α are obtained from Figure 2 at constant values of ϕ until the relationship of equation (11a) is satisfied. These calculations resemble the conventional dew point calculations and are no more difficult. However, there is a solution of ϕ corresponding to each component which is less than the relative volatility of this component and greater than the relative volatility of the next less volatile component. For each solution of ϕ there is also a value of x°_1 which corresponds to a pinch point of constant composition although only one of these points will be real and the others will occur at some points of negative concentrations. This does not interfere with the application of the equations which are exact algebraic derivations and require only that the terms be handled taking into account their proper signs.

TABLE 1
Calculation of Values of ϕ and λ for the Fractionating Section at $R + 1 = 2.78$

	x_p	α	TRIAL I $\phi_1 = 3.2$ G	TRIAL I $\frac{x_p}{G}$	TRIAL II $\phi_1 = 3.3$ G	TRIAL II $\frac{x_p}{G}$	TRIAL III $\phi_1 = 3.29$ G	TRIAL III $\frac{x_p}{G}$	$x^\circ_1 =$ ϕ	$\frac{x_p}{R\alpha G}$	x°
A	.60	4	.20	3.00	.175	3.43	.177	3.39		1.565	8.84
B	.39	2	—.60	—.65	—.65	—.60	—.645	—.605		—.560	.87
C	.01	1	—2.2	—.00	—2.3	—.00	—2.29	—.005		—.005	.00
D	.00	.5					—5.58				
				2.35		2.83		2.78		1.000	9.71
											$\lambda_1 = .103$
			$\phi_2 = 1.5$		$\phi_2 = 1.6$		$\phi_2 = 1.57$		x°_2		
A	.60	4	.625	.96	.60	1.00	.608	.987		.218	.36
B	.39	2	.25	1.56	.20	1.95	.215	1.813		.800	3.72
C	.01	1	—.50	—.02	—.60	—.017	—.57	—.018		—.016	.03
D	.00	.5					—2.14				
				2.50		2.933		2.782		1.002	4.11
											$\lambda_2 = .243$
			$\phi_3 = 1.00$		$\phi_3 = .99167$				x°_3		
A	.60	4	.75	.80	.75	.80				.112	.15
B	.39	2	.50	.78	.50	.78				.217	.43
C	.01	1		1.20	.00833	1.20				.670	80.4
D	.00	.5	— 1.0		—1.0						
				2.78		2.78				.999	80.98
											$\lambda_3 = .01235$
			$\phi_4 = .500$						x°_4		
A	.60	4	.875	.685						.048	.05
B	.39	2	.75	.520						.073	.10
C	.01	1	.50	.020						.006	.00
D	.00	.5	.00	1.555						.873	
				2.78						1.000	
											$\lambda_4 =$

Equations (12A) (12B) etc., can be most readily evaluated in terms of G by rearranging to the following form

$$x^\circ_{A1} = \frac{\phi_1}{R\alpha_A} \quad \frac{x_{AP}}{G_{A1}} \quad (15a)$$

and

$$x^\circ_{B1} = \frac{\phi_1}{R\alpha_B} \quad \frac{x_{BP}}{G_{B1}} \quad (15b)$$

etc.

Finally equation (13) can also be readily evaluated by introducing the function G

$$\lambda_1 = \frac{1}{\sum \frac{x^\circ_1}{G_1}} \quad (16)$$

where this summation refers to all the components present in the column.

All of these calculations can be carried out by a single tabulation as shown in Table 1. By assuming a value of ϕ and reading the values of G from Figure 2, it is possible to solve equation (11a) by repeating this operation until the sum of the $\frac{x_P}{G}$ column is equal to $(R + 1)$. This can be done by noting that the first three solutions lie between α_A and α_B, α_B, and α_C and α_C and α_D and also that the concentration of component D in the overhead is so small that the values of $\frac{x_P}{G}$ for component D are negligible in these cases. If subsequent calculations indicate that it must be considered, these calculations can be finally adjusted to give an exact solution but this would only be necessary if α_D is very close to α_C and, even in this case, it would affect only the calculation of ϕ_3. It can also be seen that ϕ_3 is very nearly equal to α_C or unity and by using the values of G at $\phi = 1$, the value of $\frac{x_P}{G}$ for component C is determined by differences as 1.20 for which the value of G must be .00833. Noting that the slope of the line on Figure 2 for $\alpha = 1$ is -1, this value of G will be obtained when $\phi_3 = 1 + (.00833)(-1)$ or .99167. This is not apparent on Figure 2a and must be borne in mind separately when using this type of graph. Similarly ϕ_4 must be substantially equal to α_D or .500 and by subtracting the values of $\frac{x_P}{G}$ calculated at this value of ϕ it is possible to obtain $\frac{x_P}{G}$ for component D also by difference.

The values of x° are calculated by the equations

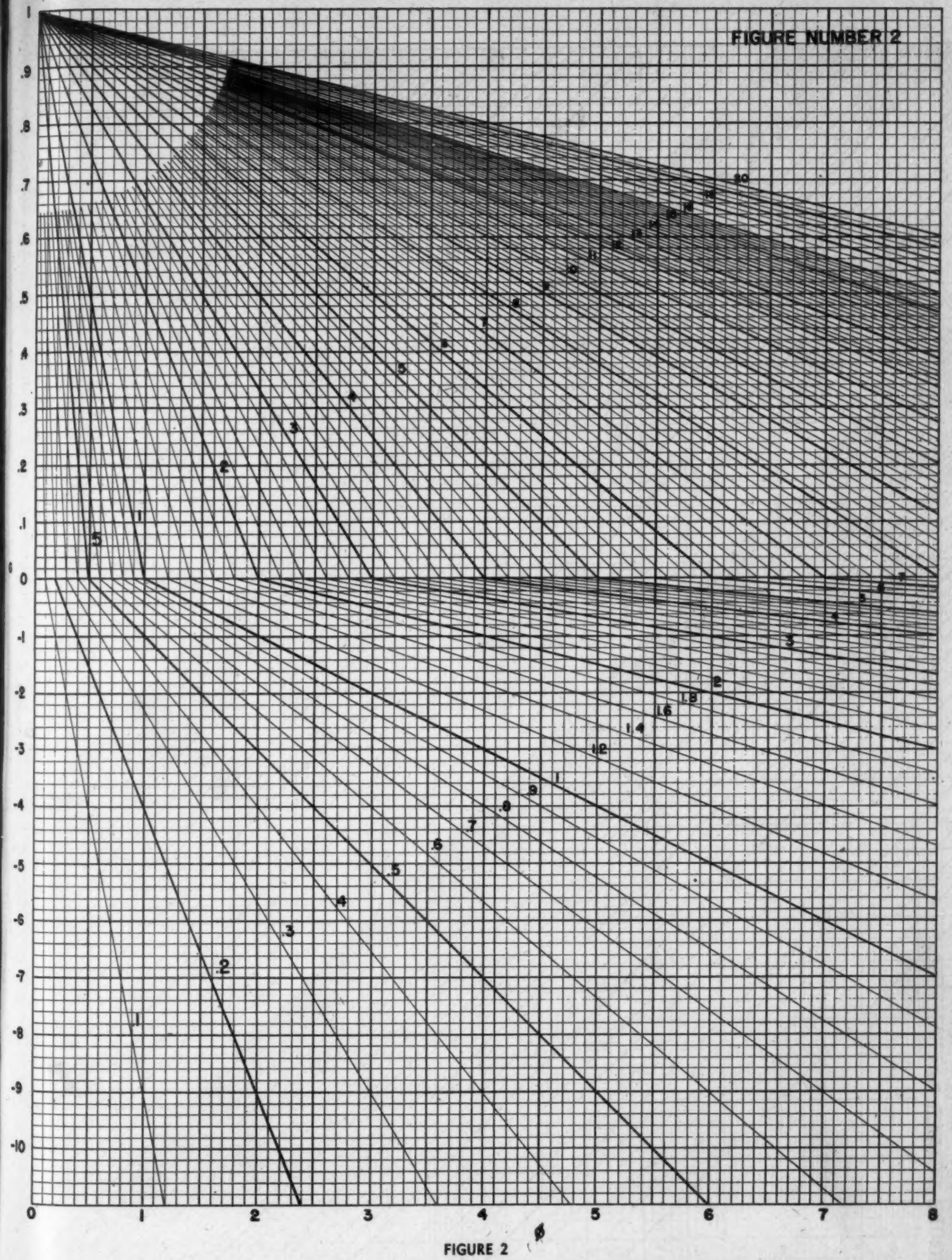

FIGURE 2

(15a)(15b) etc., noting that $\frac{x_P}{G}$ is already given the table. Thus it is possible to complete the calculation for ϕ_4 even though neither x_P or G has as yet been determined, for component D. It should also be noted that the sum of these pinch concentrations must total unity within the limits of slide rule calculation as shown and if any significant discrepancy is found, it indicates an error in the determination of ϕ. This serves as a check on the calculations up to this point.

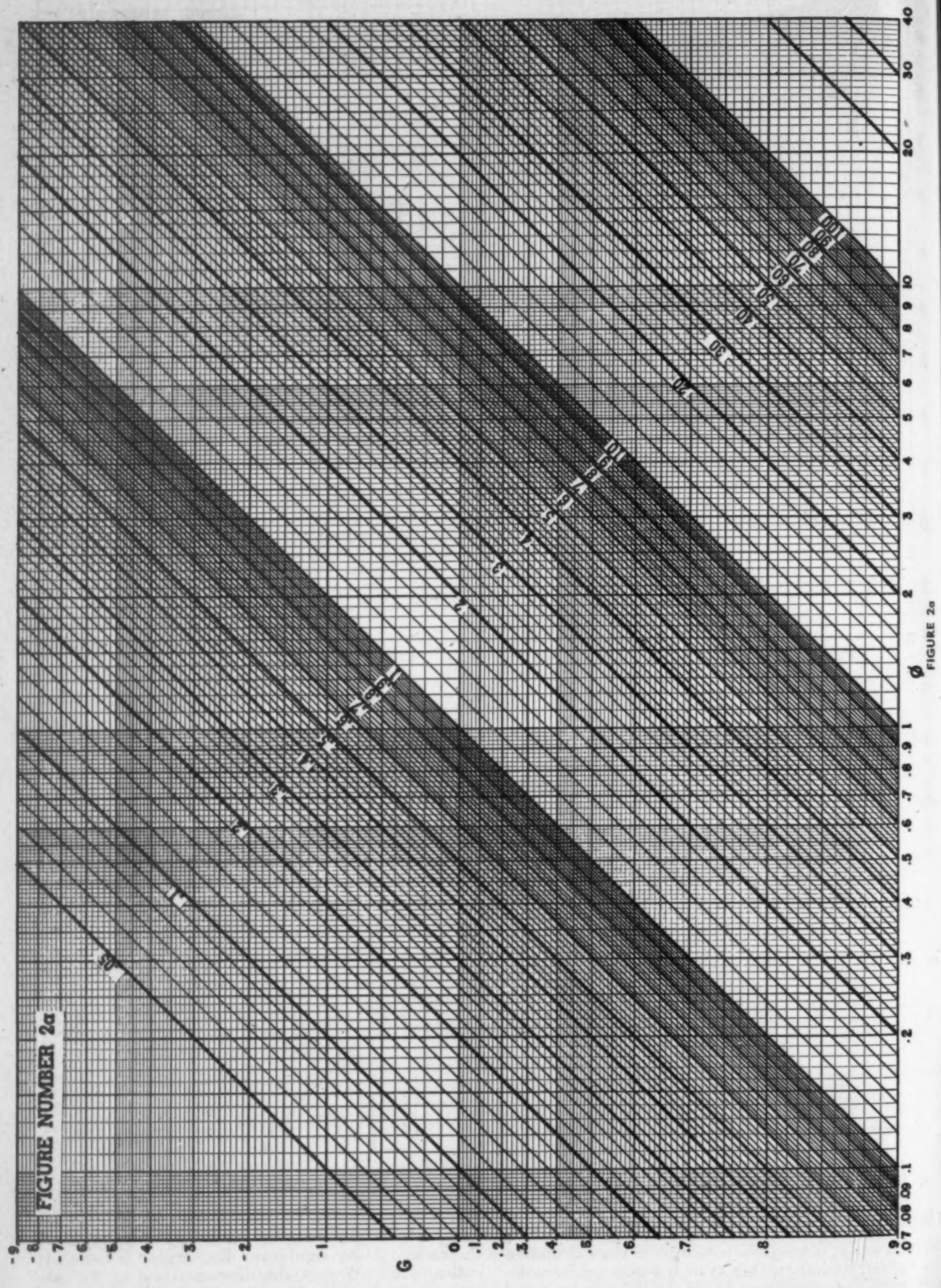
FIGURE NUMBER 2a
FIGURE 2a
G
Ø

The last column of the table gives $\frac{x^\circ}{G}$ values which are added for each value of ϕ and the value of λ obtained as the reciprocal of this sum for each value ϕ. The value of λ_4 must temporarily remain undetermined until a matching of the components is made at the feed tray.

A similar set of calculations is then made for the stripping section of the column using the equations derived by Underwood[23] similar to those for the fractionating section. The values of ϕ are then solutions of the equation

$$\sum \frac{x'_B}{G} = -S \qquad (17)$$

where x'_B is the concentration of each component in the bottoms product

S = reboil ratio, i.e., ratio of stripping section vapor to bottoms product.

In this case, the solutions are all larger than the particular values of α for the different components and smaller than the next more volatile component.

The pinch compositions for each solution of ϕ are then calculated by the equations

$$x'^\circ_{A1} = \frac{\phi_1}{(S+1)\alpha_A} \quad \frac{x'_{AB}}{G_{A1}} \qquad (18a)$$

$$x'^\circ_{B1} = \frac{\phi_1}{(S+1)\alpha_B} \quad \frac{x'_{BB}}{G_{B1}} \qquad (18b)$$

etc.

where the terms have the same significance as in the previous equations (15a) and (15b) and the prime denotes concentrations in the stripping section.

The values of λ are also given by the equation similar to equation (16) expressed in terms of concentrations in the stripping section

$$\lambda_1 = \frac{1}{\sum \frac{x'^\circ_1}{-G_1}} \qquad (19)$$

and the concentrations of the components on the mth tray in the stripping section are given by the equations similar to equations (10a) (10b) etc. but with $1/\phi$ replaced with ϕ

$$x'_{Am} = \frac{\sum_{i=1}^{i=k} \lambda_i \phi^{m_i} x'^\circ_{A1}}{\sum_{i=1}^{i=k} \lambda_i \phi^{m_i}} \qquad (20a)$$

$$x'_{Bm} = \frac{\sum_{i=1}^{i=k} \lambda_i \phi^{m_i} x'^\circ_{B1}}{\sum_{i=1}^{i=k} \lambda_i \phi^{m_i}} \qquad (20b)$$

etc.

The values of λ and ϕ can be calculated from the same general form as shown in Table 1 with overhead product concentrations replaced with the bottom concentrations and the notation of primes on all concentrations to indicate that they apply to the stripping section instead of the fractionating section.

The mechanism for the calculations given in Table 2 is similar to that for Table 1. With total liquid feed so that the stripping section vapor is the same as the fractionating section vapor $S = \frac{(1.78+1)(50)}{50} = 2.78$. Thus, a trial and error procedure is used to establish the value of ϕ when the sum of the values of $\frac{x'_B}{G}$ is

TABLE 2
Calculation of Values of ϕ and λ for the Stripping Section at S — 2.78

	x'_B	α	TRIAL I ϕ_1=4.00 G	TRIAL I x'_B/G	TRIAL II ϕ_1= G	TRIAL II x'_B/G	TRIAL III ϕ_1= G	TRIAL III x'_B/G	$x'^\circ = -\phi\ x'_B/(S+1)$	x'°/aG	$x'^\circ/-G$
A	.00	4	—.000	—2.554					.675		
B	.01	2	—1	—.010					.005		.00
C	.39	1	—3	—.130					.138		.05
D	.60	.5	—7	—.086					.182		.03
									1.000		
				—2.780						λ_1=	
			ϕ_2=2.00		ϕ_2=2.00912				x'°_2		
A	.00	4	.50	.00	.50	.00			.000		.00
B	.01	2	—.00	—2.19	—.00456	—2.19			.582	127.5	
C	.39	1	—1.0	—.39	—1.0	—.39			.207		.21
D	.60	.5	—3.0	—.20	—3.0	—.20			.212		.07
				—2.78		—2.78			1.001	127.78	
										λ_2=.00784	
			ϕ_3=1.1		ϕ_3=1.18		ϕ_3=1.165		x'°_3		
A	.00	4	.725	.00	.705	.00	.71	.00	.000		.00
B	.01	2	.45	.02	.41	.02	.42	.02	—.003		.01
C	.39	1	—.10	—3.90	—.18	—2.17	—.165	—2.36	.727		4.41
D	.60	.5	—1.2	—.50	—1.36	—.44	—1.33	—.45	.276		.21
				—4.38		—2.59		—2.79	1.000		4.63
											λ_3=.216
			ϕ_4=.55		ϕ_4=.58		ϕ_4=.581		x'°_4		
A	.00	4	.86	.00	.85	.00	.85	.00	.000		.00
B	.01	2	.72	.01	.70	.01	.70	.01	—.001		.00
C	.39	1	.45	.87	.42	.93	.42	.93	—.143		.34
D	.60	.5	—.1	—6.00	—.16	—3.75	—.162	—3.71	1.140		7.04
				—5.12		—2.82		—2.78	.996		7.38
											λ_4=.1355

—2.78. This applies to ϕ_3 and ϕ_4 in the table. The value of ϕ_2 is obtained by noting that it is very nearly equal to α_B and from the values of G at $\phi_2 = 2$ the exact value of G is obtained by difference. From Figure 2, it can also be noted that the slope of the line for $\alpha = 2$ is $-\frac{1}{2}$ and the exact value of ϕ is $2 + .00456 \times 2$ or 2.00912. This particular value does not alter the $\frac{x'_B}{G}$ calculated at $\phi = 2.00$ for the other components, but if such were the case, the calculations could easily be revised to the new value of ϕ and a more accurate evaluation of this function could be made.

Similarly, the value of ϕ_1 is very nearly equal to α_B, and using the values of G at $\phi = 4.00$, it is possible to determine $\frac{x'_B}{G}$ for component A by difference. The pinch compositions are calculated from equations (18a) (18b) etc., and the values of λ are calculated according to equation (19). It will be seen that the values of λ are the negative of those originally derived by Underwood,[23] but since these multipliers appear in all terms of the numerator and denominator of equations (20a) (20b) etc., the signs cancel out. The present definition gives positive numbers which are more convenient to handle.

When all the values of ϕ and λ have been determined for both the fractionating section and stripping section, it is possible to calculate the feed plate composition that corresponds to the optimum ratio of the key components. A study of the calculations indicates that the major factors in this ratio are the values of ϕ corresponding to the two key components or in this case ϕ_2 and ϕ_3, and the terms involving other values of ϕ may be neglected for the first estimation of the number of trays required. Thus the value of n which satisfies the following equation can be determined

$$r = \frac{\frac{\lambda_2 x^\circ_{B2}}{\phi^{n_2}} + \frac{\lambda_3 x^\circ_{B3}}{\phi^{n_3}}}{\frac{\lambda_2 x^\circ_{C2}}{\phi^{n_2}} + \frac{\lambda_3 x^\circ_{C3}}{\phi^{n_3}}} \qquad (21)$$

and by calculation of these four values in Table 3, it is possible to immediately estimate the number of fractionating trays and at $n = 7.7$, it will be found that $r = 1.01$. Similarly, by considering only the major factors in the stripping section in the following equation

$$r = \frac{\lambda_s \phi^m{}_s x'^o{}_{Bs} + \lambda_s \phi^m{}_s x'^o{}_{Bs}}{\lambda_s \phi^m{}_s x'^o{}_{Cs} + \lambda_s \phi^m{}_s x'^o{}_{Cs}} \qquad (22)$$

the number of stripping trays will be estimated between $m = 7.4$ where $r = .987$ and $m = 7.5$ where $r = 1.04$.

A first trial at $n = 7.7$ and $m = 7.35$ will indicate that the effect of components A and D in the feed tray liquid is such that the actual ratio of keys for the fractionating section is greater than the ratio for the stripping section, and consequently, additional trays would be required to match out exactly at the desired value r.

The first trial also allows an accurate evaluation of the effect of these components on the ratio of the keys and the second trial at $n = 7.95$ and $m = 7.55$ is shown in Table 3. A table of this form allows a rapid calculation of feed tray composition at different value of n and m and is very useful in estimating trends and the effect of the other components on the concentrations of the keys so that it will be immediately apparent before all the calculations are completed whether the components are going to match out. Also the preliminary trials can be made by equations (21) and (22) by merely filling in the four appropriate values required by these equations and when the desired ratios are obtained, the balance of the table can be calculated. Thus, the use of Underwood's equations in a table of this form represents a tremendous saving of time over the conventional trial and error tray calculations which are exceedingly difficult to carry out for fractional numbers of trays and requires plotting several curves to finally match the components.

From the vales of ϕ and λ in Tables 1 and 2, it is possible to complete the first three lines of the fractionating part of Table 3 and the last three lines of the stripping part of Table 3. By neglecting the missing terms, it is then possible to calculate the ratio of the component A (lighter than the keys) to the key components in the fractionating section and then introduce the necessary amount of A in the stripping section calculation to give this ratio. Thus, the first calculation on this table would give

$$\frac{.00295}{.01699} \times 1.688 = .293$$

to be inserted as the value of $\lambda_1 \phi_1{}^m x'^o{}_{A_1}$ in the table. It is then possible to calculate the other values by the ratio of their pinch compositions, $x'^o{}_1$, in Table 2. Thus, the ratio of component D to the keys can be calculated which can be used in the last line of the first part of Table 3. These values are all shown in parenthesis in the table. The second set of figures are then made in a similar manner starting with the new value calculated as

$$\frac{.00327}{.00875 + .00877} \times (.879 + .871) = .327$$

and the other figures in the table correspond to this second cycle of calculations. If a third cycle of calculations were made, the new starting value would be .329 as compared to .327 and this would not change

TABLE 3

Calculation of Feed Plate Compositions

	ϕ	λ	$\frac{\phi^n}{\phi^m}$ or $\lambda\phi^m$	$\frac{\lambda}{\phi^n}$ or $\lambda\phi = x'^o{}_A$	$\frac{\lambda}{\phi^n} x^o{}_B$ or $\lambda\phi = x'^o{}_B$	$\frac{\lambda}{\phi^n} x^o{}_C$ or $\lambda\phi = x'^o{}_C$	$\frac{\lambda}{\phi^n} x^o{}_D$ or $\lambda\phi = x'^o{}_D$
				n = 7.95			
1	3.29	.103	12,900	.0000080	.00000	.00000	.00000
2	1.57	.243	36.0	.00675	.00147	.00540	−.00011	.00000
3	.99167	.01235	.936	.01319	.00148	.00286	.00884	.00000
4	.500		.0040	.00687	(.00032)	(.00049)	(.00004)	(.0058)
					.00033	.00050	.00004	.00600
	Concentrations			.02681	(.00327) .00328	(.00875) .00876	(.00877) .00877	.00600
					.122	.327	.327	.224
				m = 7.55				
1	4.00		39,000	.488	(.293)	(.002)	(.060)	(.079)
2	2.00912	.00784	193	1.510	.327	.002	.067	.088
3	1.165	.216	3.17	.685	.000	.879	.313	.320
4	.581	.1355	.0165	.0022	.000	−.002	.498	.189
					.000	.000	.000	.003
				2.685	.327	(.879) .879	(.871) .878	(.591) .600
	Concentrations				.122	.327	.327	.224

any of the other figures in the table so the concentrations of A and D are balanced with the keys, and it is then necessary to compare the actual concentrations of all components. It should be noted that the values of $\frac{\lambda_4}{\phi^n{}_4}$ and $\lambda_4 \phi^m{}_4$ are the sums of the concentration terms in the same line. The concentrations all agree perfectly and the ratio of the keys is 1.00 which is exactly that of the optimum feed tray so the column design is exact. In actual practice a perfect match such as this would not be necessary and, if all the concentrations agreed within 1 percent, the solution would be sufficiently accurate. If the ratio of the keys in one section was the optimum but the other section did not agree with it more or less trays would have to be added to the other section of the column and the following general rule may be employed. In the fractionating section the ratio of light to heavy key decreases with number of trays and the same ratio in the stripping section increases with the number of trays. It is possible to match all the component at almost any point in the column but unless the ratio of the keys in the match is nearly equal to the calculated optimum ratio, the calculations should be repeated. With practice these calculations can be made very rapidly and an entirely exact design of a column can be made which would require a much larger amount of time by the tray-to-tray method of calculations.

It should be noted that with the feed introduced on the tray, 7.95 below the top, 6.95 fractionating trays are required and 7.55 stripping trays. The previous tray-to-tray calculations gave 7.0 fractionating trays and 7.3 stripping trays and the discrepancy is probably due to a slight cumulative error in the tray-to-tray calculations. The trays calculated by Underwood's equation should be more exact because slide rule errors are involved only in a single equation, even though this is a relatively complicated equation. The situation is analogous to raising a number to the tenth power by multiplying it ten times on a slide rule as compared to obtaining the answer from a single equation. Thus, the equations of Underwood can give a simpler and more precise answer than the previous tray-to-tray calculations.

By calculating the previously undetermined terms in the Tables 1 and 2, it can be found that in the

fractionating section $\lambda_4 = .0040 \times .00687 = .0000275$ and $\frac{x^o}{G} = 36300$ and $G_{D_4} = .000024$. Thus, $\phi_4 = .499988$ and $x_{DP} = .000037$. By using these values to repeat the previous calculations in Table 1, no effect will be observed on the calculations. If such an effect were noticeable, it would be necessary to adjust the calculations accordingly. Such adjustment is unusual and would always be small so that it would never have to be repeated more than once to obtain a more accurate set of figures.

Similarly for the stripping section it will be found that $\lambda_1 = .0000125$, $G_{A_1} = .0000084$, $\phi_1 = 4.000034$ and $x'_{AB} = .000021$ and consequently they will have no effect on the calculations in Table 2. The calculations also give an evaluation of the concentrations of all the components in the overhead and in the bottoms although it is apparent that the quantity is negligible and would not affect the overall material balance.

Underwood[21, 22] has also derived a method for determining the minimum reflux ratio which works out very easily by the use of G vs ϕ chart (Figure 2). The basic equation is

$$\sum \frac{X_F}{G} = 1 - q \qquad (23)$$

where the summation refers to all components in the feed and q is the fraction of the feed which is liquid or, in terms of enthalpy, it is the heat required to vaporize the feed completely divided by the latent heat of the feed. For a total liquid feed at the feed tray temperature $q = 1$ and for a total vapor feed at the feed tray temperature $q = 0$.

The values of G for the solution of ϕ between the relative volatilities of the key components are substituted in equation (11a) to give the minimum reflux ratio

$$\sum \frac{x_F}{G} = R + 1 \qquad (11a)$$

This calculation is given in Table 4 in which the three trials required to evaluate ϕ are shown. The exact minimum reflux requires that the value of ϕ be determined carefully because the sum of $\frac{X_F}{G}$ is composed of terms of opposite signs and varies to a much less degree than all the positive terms. The sum of $\frac{x_F}{G}$ consists primarily of the positive terms and also since the reflux ratio is finally determined by subtracting unity from the sum, the minimum reflux ratio is very sensitive to the value of ϕ. However by careful calculations the minimum reflux ratio can be evaluated with a higher degree of accuracy than previous methods. The calculated minimum reflux ratio is .885 and this is in agreement with that previously determined by tray-to-tray calculations in which .89 was found to give a satisfactory match although the calculated minimum reflux ratio is more accurate and will probably give a better match of the component at the feed tray. The reflux ratio calculated by equation (3) was .912 and, while this is somewhat easier to use, especially when the binary equilibrium curve of the keys is available, the more exact value given by Underwood's equation completely justifies the extra work required to determine it. It should be noted that part of the work in applying equation (3); namely, the determination of x_1, is essential for the use of equation (7) in locating the optimum feed tray. Thus equation (3) can be used as a rapid check of the calculations because the both methods should give fairly good agreement.

TABLE 4
Calculation of Minimum Reflux by Underwood's Method

			TRIAL I $\phi = 1.5$		TRIAL II $\phi = 1.25$		TRIAL III $\phi = 1.257$			
	X	α	G	$\frac{X_f}{G}$	G	$\frac{X_f}{G}$	G	$\frac{X_f}{G}$	x_p	$\frac{x_p}{G}$
A	.30	4	.625	.48	.687	.437	.686	.438	.60	.875
B	.20	2	.25	.80	.375	.534	.372	.538	.39	1.049
C	.20	1	−.5	−.40	−.250	−.80	−.257	−.778	.01	−.039
D	.30	.5	−2.0	−.15	−1.5	−.20	−1.514	−.198		
				+.73		−.029		.000		1.885

Minimum Reflux Ratio = .885

It is also interesting to note that when the pseudo reflux ratio is calculated from equation (6) using the exact minimum ratio, it gives an agreement of about 5 percent between the trays required by the McCabe-Thiele method at this reflux ratio as compared to the trays calculated by the method of Underwood which is considered more accurate than the actual tray-to-tray calculations.

C) Application to Other Systems

In the present method two basic assumptions are made which are used in most previous approaches to the solution of multicomponent distillation problems. The first assumption is that of constant relative volatility and the second is that of constant molal downflow or in other words constant L/V ratio in the fractionating section and in the stripping section of the column.

For the case of varying relative volatility it is possible to use an average value at the temperature of the feed plate. It is difficult to predict this temperature in advance however, the temperature of the feed plate based on the vapor liquid equilibria data of the keys and the pseudo reflux ratio is readily determined and can be used satisfactorily in the present method.[17] This procedure is most accurate in the minimum reflux ratio calculation because under this condition the region close to the feed establishes the minimum reflux ratio even though the relative volatilities vary above and below the feed.

In general with mixtures of hydrocarbons the relative volatility increases at lower temperatures and consequently the use of the average value in the equations of Underwood would give a slightly larger number of trays above the feed and a slightly smaller number of trays below the feed than actually required. Thus the total number will be substantially correct but the feed plate location will be somewhat lower than required. A more accurate calculation would be obtained by using a different average relative volatility for the fractionating and stripping sections in these equations.

However, the errors introduced by the use of average relative volatilities for systems of hydrocarbons are usually negligible compared to the errors introduced by the assumption of constant molal downflow in the two sections of the column. The completely rigorous method of tray to tray calculations involving heat balances on every tray to establish the vapor and liquid quantities on each tray was described by Sorel.[12] Establishing heat balances on each tray is a time consuming operation however recently charts have been developed which greatly simplify this work.[14, 15, 16]

The present method for determining the number of trays in a column could be reliably used for these systems by calculating the reboiler duty from a heat balance below the feed tray at the operating reflux ratio or by calculating the condenser duty by a heat balance above the feed tray. In either case the other one of these duties would be obtained from the overall heat balance on the column. By establishing conditions at the feed tray in this manner it is possible to insure the operability of the column because the tendency to pinch out and prevent the desired separation from being achieved is greatest at this point. When the feed tray conditions are properly established the variation of the L/V ratio above and below the feed produces a curved operating line which affects only the total number of trays required but will not make the separation impossible.

Considerable difficulty is usually encountered with these calculations due to the condition of the feed because a totally liquid feed will be much below the feed tray temperature and a totally vapor feed will be much above the feed tray temperature and may even be above the temperature of the bottom of the column in the case of a large amount of material heavier than the keys. A partially vaporized feed is between these limits and the liquid can act as either a sub-cooled or super-heated feed producing either less or more vapor in the column. Also the flash vapor composition bears no relation to the feed tray vapor as in the case of a binary system. Thus considering the equilibrium flash of the feed produces a single set of conditions at the feed tray which is inconsistent with all other conditions in the column and the simplest way to handle it, is to consider the feed as a liquid which enters the column with an additional amount of heat to vaporize a certain amount of the liquid in the column instead of outside the column. Thus this additional vapor which is produced by the heat content of feed will be of the same composition of the feed plate vapor and a completely consistent set of conditions will be calculated in the tower. In theory this would amount to introducing the vapor so it passes through the feed tray liquid instead of directly to tray above the feed as it usually does in practice. This assumption will produce no significant errors in the calculation and will prevent the false impressions gained by considering the flash of the feed because the actual effect of the feed on the operation of the column bears no relation to this equilibrium flash of the feed. The design of the column is dependent solely on the actually enthalpy of the feed and a preliminary heat balance should be made to estimate its effect on the relative quantities of vapor and liquid in the column when there is considerable doubt about it. However its effect will be accurately evaluated in the final calculations so it is not essential to evaluate it exactly at the start.

When the feed plate composition has been established by the present method it is possible to carry out one set of complete tray-to-tray calculations by the Sorel method to determine the exact number of trays required above and below the feed when heat balances are considered. This is a great improvement over the previous methods in which the trial and error procedure of matching the components at the feed tray is further complicated by the heat balances at the feed tray. Also in this case a single final set of tray-calculations and heat balances would be required which could also take into account the varying relative volatility in the column. This would be the most reliable design that could be calculated.

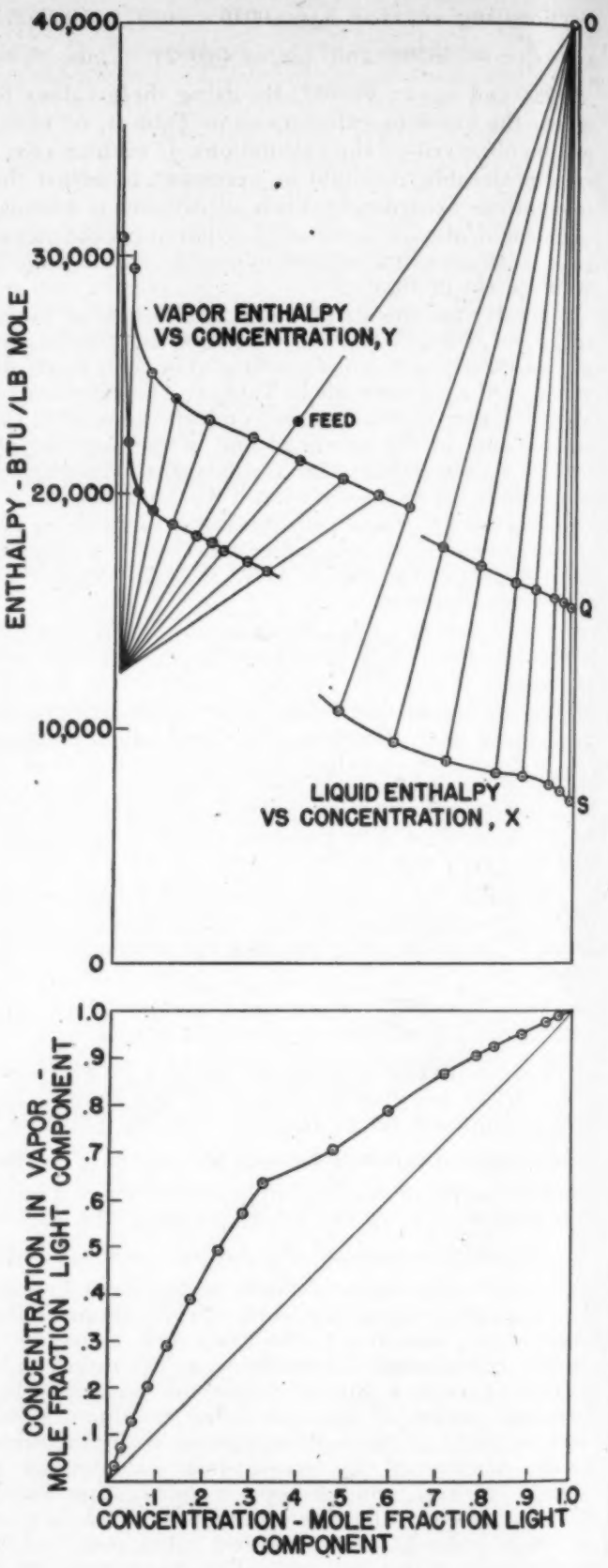

FIGURE 3

If this final complete set of calculations did not give the desired number of trays as; for example, when a new feed stock is to be processed in existing equipment or in an existing design which is to handle more than one feed, it is possible to extend the principles applied by Jenny and Cicalese[9] to the McCabe-Thiele method[12] to include the Ponchon[11, 12] method. The method of Jenny and Cicalese consists of lump-

ing all the components lighter than the light key with this component and all the components heavier than the heavy key with this component to obtain a two-component mixture which can be handled on the McCabe-Thiele diagram. The method requires one complete set of tray to tray calculations so the equilibrium curve can be determined and the trays required at other reflux ratios are stepped off graphically. The method is not exact because the equilibrium curve determined in this manner actually varies with the reflux ratio and it is most reliable for small changes in the reflux ratio. If the preliminary design was close and it should have been within 15 percent if any of the approximation methods were used, a revised design by this method to give the desired number of trays should be completely sound.

The previous method is based on constant molal downflow and subject to all the limitations of this assumption. However, if a Ponchon diagram is constructed from a set of heat balances and based on the previous definition of composition in terms of a binary system it is possible to locate a new operating point at a different reflux ratio and carry out the graphical step wise calculations based on the combined Ponchon diagram and the equilibrium curve. Figure 3 shows a Ponchon diagram constructed from a set of tray calculations and heat balances on a debutanizer. These were made using the K-value and enthalpy charts previously mentioned.[14] The calculations were made for the stripping section and the feed was introduced on the eighth tray above the bottom. According to the previous criterion the optimum feed tray was between the eighth and the ninth tray above the bottom. It can be seen that the optimum feed tray is several trays above that located by the usual method on the Ponchon diagram. This is common to all cases with a very large percentage of components heavier than the keys and has been discussed in a previous paper.[18]

It is also interesting to note that the enthalpy of the liquid is such that it will produce more than its own number of moles of vapor in the column even though the equilibrium flash on the feed gives less than 50 percent vapor. The enthalpy curves for the fractionating and stripping sections are discontinuous and at different reflux ratios the graphical tray calculations should be started close to the original feed tray to minimize extrapolation of these curves. Also the actual calculations and the graphical calculations at different reflux ratios should be started at the top and bottom so that the fractional number of trays can be estimated more reliably in the middle parts of the fractionating and stripping sections in which the variations are more uniform. It is apparent that estimation of fractional trays at the bottom is difficult because of the peculiar shape of the enthalpy curves and the same effect is present in the fractionating section when a large concentration of components lighter than the keys is involved.

The calculations shown were carried out for an external reflux ratio of 3:1 and thus the ratio of the length of line OQ to line QS is 3. The focal points for other reflux ratios could be located accordingly. Also the line from O through F locates the focal point for the stripping section at the bottom concentration. The equilibrium curve proposed by Jenny and Cicalese is given in the lower section of the figure to facilitate the location of the vapor and liquid concentrations which are in equilibrium. This is a generally recommended procedure for applying the Ponchon method[12] because it simplifies the construction.

The Ponchon diagram can also be constructed using enthalpy values in Btu / lb. but in this case the concentrations must be in weight fractions and this compresses the stripping section into a much smaller range of concentrations and thus requires an additional enlargement of the diagram in this region for accurate graphical calculations.

The over-all accuracy of the graphical method is of course dependent upon the scale used and its inherent accuracy is about the same as that of Jenny and Cicalese which neglects heat balances on the column.

It is based on the essentially same assumption; namely that the equilibrium curve does not change with reflux ratio and in this particular case that the enthalpy curves do not change with reflux ratio. The method is most reliable at reflux ratios close to that upon which the equilibrium enthalpy curves were based but if the preliminary reflux ratio were chosen as previously described the final reflux ratio will not be sufficiently different to introduce any appreciable errors by the graphical procedure.

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Economic Distillation Design

A Study of the Comparative Importance Of the Distillation Variables

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IN THE study of a distillation system there are three principal types of variables as follows:

1) The separations obtained between the components in the overhead and bottoms product streams.

2) The heat effects of the condenser, reboiler, and feed preheater, and the resultant amounts of the countercurrent liquid and vapor streams.

3) The numbers of vapor-liquid contacts or "plates" above and below the feed inlet.

The plate ordinarily used in the calculations is the "equilibrium" or "theoretical" plate for which the leaving liquid and vapor streams are considered to be in equilibrium. Numbers of actual and theoretical plates are connected by simple ratios, or "plate efficiencies," based on experience.

In the commercial practice of distillation the usual object is the simple making of money. It would seem, then, that a study of the comparative importance of the above listed variables could be based on economic considerations. A gasoline stabilizer might be considered to be a typical distillation installation. For one such unit, and for the particular system of accounting used, the total cost of the stabilizing operation for one year was about $65,000. The cost of the reboiler heat and condenser cooling for one year for the unit was about $17,000, and this includes the cost of maintaining the heat transfer equipment. The original cost of the 24-tray column installed was about $9000, and the combined maintenance and depreciation for one year averages $1000. It can be estimated that the cost of one tray and the section of tower it occupies is about $35 per year.

The separation of the butanes and lighter, the reason for the distillation, is certainly the most important of the variables and is worth more than the $65,000 total cost for the operation. And on the basis of yearly cost the heat and cooling requirements necessary to obtain these separations are easily second in importance ($17,000). The number of plates variable is seen to be a very poor third in cost per year. And yet a major concern of writers of distillation theory has been to determine the numbers of plates required for an operation when the separations and heat duties are fixed. Some of the methods presented are of astonishing complexity, and the answers obtained are expressed in fractions of a plate.

The basic economic unreality of the number of plates approach to distillation design has led to a general discrediting of the more involved plate-to-plate methods and a demand by "practical" engineers for a more fundamental design procedure. One rule that has been applied is: The economic number of plates for a separation is twice the minimum number as calculated with Fenske's equation[3]. However rough and incomplete this rule is it is more likely to lead to a more economic design than the chance of arbitrarily selecting a "reasonable" reflux and elaborately calculating the resultant required number of plates.

It is true that it can be difficult or impossible to obtain a desired separation without sufficient plates. But also it would be impossible to obtain the separation without a reflux pump. And yet the existence of a reflux pump is not considered to be a distillation variable. It is the cost of insuring that a facility is present rather than the consequences of its not being present that determines its comparative importance in an economic analysis. The number of plates should be "adequate" and some means of determining such a number should be available. Similarly the tower should have sufficient cross-sectional area, and methods for finding this area have been given[4]. Both the number of plates and the tower diameter are construction features, and the accuracy of their determination would seem to be of about equal importance economically.

RECENTLY a general method for the determination of economic optimums was presented by the writer[2] and was applied to the operating conditions of plants already in existence. The same mathematical system can be used for the first design of a plant, and is here applied to the design of a distillation operation. The nomenclature used is as follows:

a, b, c . . . e—Design variables measured as units
A, B, C . . . E—Cost or value in dollars per time of an incremental unit of the corresponding design variable.

These cost terms are positive for product values and other plant gains, and are negative for necessary costs such as reboiler heat.

At the optimum design these terms are connected by the following relations (see Reference 2).

$$A\,(\partial a/\partial b)_{c\ldots e} + B = 0 \quad \text{(lab)}$$
$$A\,(\partial a/\partial c)_{b\ldots e} + C = 0 \quad \text{(lac)}$$
$$B\,(\partial b/\partial c)_{a,e\ldots} + E = 0 \quad \text{(lbe)}$$

etc. for any other combination of two variables.

For a system of n design variables only n-1 of these equations can be considered to be independent. The remaining relation necessary to solve for the n unknowns is the inter-relations of the design variables.

In the use of these equations to solve for a given optimum design two types of data are required. First, the complete inter-relations of the design variables for the conditions given must be known, together with the resultant values of the partial terms $(\partial a/\partial b)_{c\ldots e}$, etc. Second, the values or costs in dollars of additional units of the design variables—values of A, B, C . . . E—must be known, either as constants or as functions of a, b, c . . . e. It will be noted that these cost terms are for incremental units of the design variables and are not averaged values of total cost divided by total number of units.

In general the solution of a given system of data for an optimum design with these relations is simple and direct, although occasionally trial and error methods are required. A typical procedure is given

in the simplified example that follows.

A chemical by-product is obtained at a constant rate as a 28 mol percent solution in water. It is desired to recover this by-product, at a purity corresponding to 99.2 mol percent, as the overhead of a distillation column. One percent recovery of this chemical for one day's operation is worth $4.20. Sufficient heat pickup is available to bring this feed to the boiling point. The relative volatility for dilute water in chemical (top of column) is 1.58, and for dilute chemical in water (bottom of column) is 1.75.

The cost of one additional MM (million) Btu. of reboiler heat to the column is $0.31. This figure includes the cost of the heat itself, the cost of the condenser cooling to remove the heat, and the cost of the additional reboiler and condenser surfaces. It does not include the cost of the additional plate area required to handle the vapor.

For the completely installed column the cost of one additional plate together with the section of tower it occupies is estimated to be $96 plus $44 per square foot of area. With 18 inch plate spacing one square foot of plate area will handle a vapor load corresponding to 20 MM Btu. of reboiler heat per day. The plate efficiency for these conditions is taken at 90 percent. It is estimated that the combined depreciation and maintenance for the column for one year (300 operating days) will be one fifth of the installed cost.

It is desired to obtain the most economic design for this distillation operation.

As hydrocarbon processing plants become more complex and a wider variety of end products are manufactured, the problems of plant economics, basically simple in nature, become increasingly difficult through the greater number of variables that must be considered. Optimum efficiencies can be more readily achieved when efforts at plant improvement are expended for each of these many variables in proportion to its importance in setting overall profits.

The variables used in the solution for this optimum are defined as follows:

a — Recovery of chemical in percent of the total.
b — Reboiler heat duty in MM Btu. per day.
c — Number of equilibrium plates in the tower.

The inter-relations between these variables as calculated for the given conditions are shown on Figure 1. The product recovery is plotted vs. the reboiler heat input for towers of various numbers of plates (a vs. b at constant c). These curves were obtained by a trial and error arrangement using McCabe-Thiele[3] diagrams. A single point on this graph represents the optimum design.

The first step in determining this optimum is to express the costs of additional units of these variables (values of A, B and C) on the basis of dollars per operating day. For the product recovery variable the value given for one additional percent for one day is

$$A = 4.20$$

For the reboiler heat variable the cost per day for one MM Btu. is 0.31 plus the cost of the additional plate area required. From the data given the cost of this area (1/20 of a square foot) for an equilibrium plate is

$$(44.0/20)/(5 \times 300 \times 0.9) = \$0.00163 \text{ per day}$$

where (44.0/20) is the first cost of the area, 5 is the fraction of this cost per year, 300 is the operating days per year, and 0.9 is the plate efficiency. Then for a tower of c plates this cost becomes 0.00163c so that

$$B = 0.31 + 0.00163c$$

For the number of plates variable the total area required per plate is b/20 square feet so that the original cost of one additional plate is $96.0 + 44.0(b/20)$. Then the cost per day of an equilibrium plate becomes

$$C = \frac{96.0 + 44.0\ (b/20)}{5 \times 300 \times 0.9} = 0.071 + 0.00163b$$

With these values equation (lab) is substituted in to give

$$4.20\ (\partial a/\partial b)_c - (0.31 + 0.00163c) = 0 \qquad \text{(lab)}$$

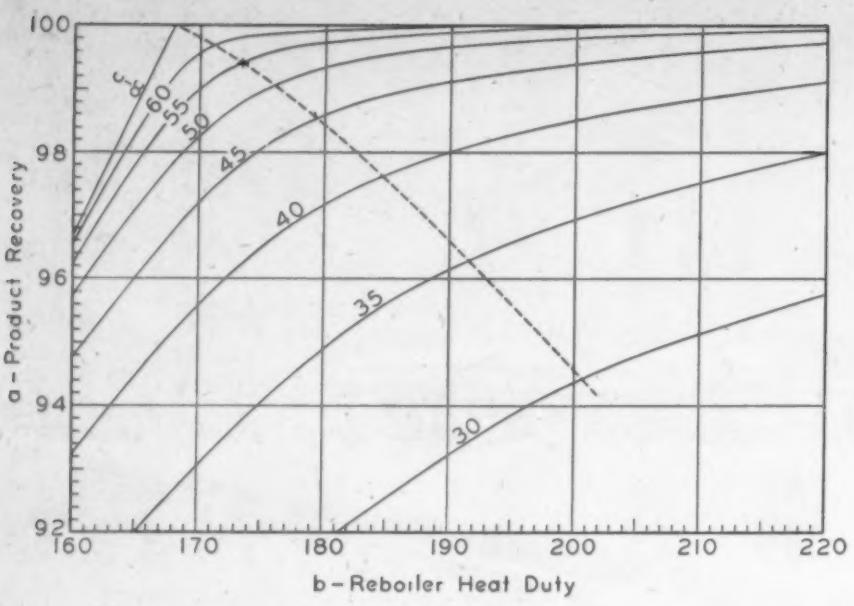

FIGURE 1—Recovery of product in percent vs. reboiler heat duty in MM Btu. per day for towers with given numbers of equilibrium plates (a vs. b at constant c).

Dashed line—Values that satisfy Equation lab. See text.

where, of course, $(\partial a/\partial b)_c$ is the slope of the curves of Figure 1. If $c = 50$ is substituted in this equation $(\partial a/\partial b)_c = 0.093$ is obtained. The point on the 50 plate curve that has this slope is located and marked. Similar points are obtained for the curves for other numbers of plates and are connected to give the heavy dashed line of Figure 1. These values satisfy equation (lab) so that the optimum design lies somewhere on this curve.

The next step is to cross plot the data of Figure 1 as percent recovery vs. number of plates for constant amounts of reboiler heat (a vs. c at constant b) as shown in Figure 2. Equation (lac) is substituted in to give

$$4.20\,(\partial a/\partial c)_b - (0.071 + 0.00163b) = 0 \qquad \text{(lac)}$$

in which $(\partial a/\partial b)_b$ is the slope of the curves of Figure 2. If $b = 180$ is substituted in this equation then $(\partial a/\partial b)_b = 0.087$ is obtained. The point at which the $b = 180$ curve has this slope is located and marked. Similar points are obtained for the curves for other amounts of reboiler heat, and these are connected to give the heavy dot-dash line of Figure 2. Values from this curve satisfy Equation lac and so the optimum design lies somewhere along this line.

Values from the dashed line of Figure 1 are cross plotted to give the dash line of Figure 2. Obviously the point at which the dot-dash and dashed lines cross satisfies both equations (lab) and (lac), and so gives the desired optimum. The values $a = 99.4$ and $c = 55$ are read from this point, and these values plotted back on Figure 1 (marked by cross) give $b = 173.5$. Then as the final economic design the tower should have 55 equilibrium plates and the reboiler heat load should be 173.5 MM Btu. per day. The resultant recovery of the chemical is 99.4 percent. The tower would have $55/0.9 = 61$ actual plates with a vapor area of $173.5/20 = 8.7$ square feet minimum. This design would give the maximum profit for the operation.

Loss from Errors for Example

At the beginning of this paper the distillation variables were compared on the basis of total cost per year. But in design work a better measure of their comparative importance would seem to be the cost of errors in the determination of their optimum values. Thus for the distillation of the example 45 equilibrium plates might have been used instead of the optimum 55. The cost per day of the plates would be decreased, but this would be more than offset by the increases in the costs of the reboiler heat and of the product loss so that a net loss as compared with the optimum would result. The actual values for this loss can be calculated as follows:

On Figure 1 the optimum operation for a tower of 45 plates is the point at which the heavy dashed line crosses the 45-plate line, and the values $a = 98.52$ and $b = 179.4$ can be read for this point. Then the value of the product lost as compared with the optimum is $(99.40\text{-}98.52) \times 4.20 = \$3.70$ per operating day. Also the cost of the additional reboiler heat is $(179.4\text{-}173.5) \times 0.31 = \$1.83$ per day. With the expression $C = 0.071 + 0.00163b$ the decrease in plate cost can be determined as $(55 \times 0.354)\text{-}(45 \times 0.363 = \$3.15$ per day. Then the net loss for a 45-plate column as compared with a 55-plate column is $3.70 + 1.83\text{-}3.15 = \$2.38$ per operating day. This and other values obtained in a similar manner are given in the following table.

TABLE 1

Loss from Error in Number of Plates

Number of Plates	40	45	50	55	60
Percent from Optimum	−27.3	−18.2	−9.1	0	+9.1
Decrease in Return (dollars per operating day)	6.34	2.38	0.68	0	0.46

Corresponding daily losses resulting from amounts of reboiler heat other than the optimum can be calculated similarly. The optimum operation at the heat duty used is read from the dot-dash curve of Figure 2. The losses per day obtained are as follows:

TABLE 2

Loss from Error in Reboiler Heat Duty

Reboiler Heat Duty (MMBtu. per day)	160	168	173.5	180	190
Percent from Optimum	−7.8	−3.2	0	+3.75	+9.5
Decrease in Return (dollars per operating day)	10.20	0.95	0	1.13	3.75

The high loss for the 160 MM Btu. operation is due to the decreased recovery of product, and an infinite number of plates would change this but little (see Figure 1).

It is evident from the values of Tables 1 and 2 that the same percent of error is much more important for the reboiler heat duty than for the number of

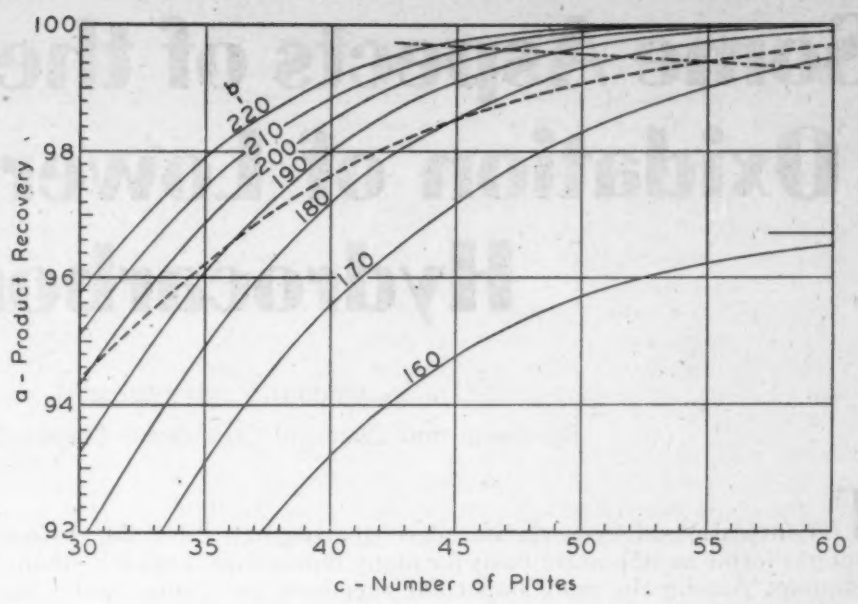

FIGURE 2—Recovery of product in percent vs. number of equilibrium plates for various amounts of reboiler heat duty in MM Btu. per day. (a vs. c at constant b.) Cross plot of Figure 1.

Dot-dash line—Values that satisfy Equation lab. See text.

Dashed line—Values that satisfy Equation lab. Cross-plot from Figure 1.

plates. Thus the effect of a 5 percent error in the numbers of plates can be determined by plotting the values of Table 1 (and others) as percent error vs. loss per operating day. For an error of —5 percent (55 x 0.95 plates) the loss is read from this curve as $0.25 per operating day, and for an error of +5 percent (55 x 1.05 plates) the loss is read as $0.18 per operating day.

The corresponding losses from 5 percent errors in the reboiler heat duty are found by plotting the values of Table 2 (and others) also as percent error vs. loss per operating day. For an error of —5 percent (173.5 x 0.95 MM Btu.) the loss is read from this curve as $3.95 per day, and for an error of +5 percent (173.5 x 1.05 MM Btu.) the loss is $1.70 per day. Evidently for 5 percent errors the losses for the reboiler heat duty are about 10 times as great as for the number of plates. It is believed that the conditions and costs given for this example are weighed in favor of rather than against the importance of the number of plates variable.

It is seen that for the particular conditions given any number of plates from 45 to 60 or more can be used without too serious consequences. On the other hand the tower heat duties are important and need to be designed and controlled closely. The tower operation should be carefully "balanced" so that the maximum separations are obtained for the heat duties used.

Discussion

A point of importance in design calculations of the type given here is that the data used should be based on most probable values and should not include oversizes to make the design "safe." An accumulation of factors of safety makes the calculated optimums meaningless. Thus in the example, if tray sizes larger and more expensive than necessary had been included in the data then fewer trays would have been calculated for the final design, and the optimum would have been missed. Once the optimum design has been determined for the most probable operation, oversizes for safety and flexibility can be added as judged necessary.

A value that is frequently difficult to estimate is the cost per operating day of equipment such as a distillation column. This value includes both maintenance and depreciation on the first cost. Many companies require a certain "pay-out" period for new construction, and also depreciate the first cost of the plant over a standard number of years for tax purposes. Neither of these bookkeeping terms has any connection with the actual corrosion or obsolescence that determines the life of the equipment. As stated above the designer must select a most probable life for the operation based on his own judgment of the factors involved.

From the high ratios of product value and of heat cost to plate cost it can be concluded that most towers should have comparatively large numbers of plates, so that the operation approaches that for a tower of infinite plates. Thus for the example given the optimum reboiler heat duty was only 3.3 percent more than would have been required for a complete recovery of the product at infinite plates. This can be compared with the rule sometimes used that the heat duties should be 20 percent more than the minimum for the given separation, the so-called "minimum reflux."

The most important decisions made in distillation design are the separations to be obtained, and these values are frequently arbitrarily set. With the additional relations between the variables given by the mathematical system described, the number of such arbitrary decisions, with their attendant probable error, can be reduced to a minimum. It is believed that an amount of judgment organized along the lines described will more probably lead to a more economic design than the same amount of judgment unorganized.

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Some Aspects of the Industrial Oxidation of Lower Aliphatic Hydrocarbons

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THE oxidation of hydrocarbons derived from natural gas forms an important basis for many industrial reactions. Among the more important chemicals derived from the direct oxidation of gaseous hydrocarbons are acetaldehyde, formaldehyde, propionaldehyde, methanol, ethanol, propanol, acetone and acetic acid. The industrial importance of direct oxidation is borne out by the fact that Cities Service Oil Company at Tallant, Okla., produces substantial amounts of methanol, formaldehyde and acetaldehyde from the direct oxidation of methane, ethane and propane.[7] Also, Celanese Corporation of America has built a $20 million plant for the partial oxidation of 15 billion cubic feet per year of natural gas.[2]

The vast quantities of hydrocarbons available in crude petroleum and in natural gas offer a source of materials for organic synthesis at relatively low cost. Such possibilities, inherent in oxidation which might convert hydrocarbons to oxygen-bearing substances of commercial value, make the study of such operations of importance. As an example of the possibilities in this field, Egloff and Schaad[3] in 1929 calculated from the amounts of the various paraffin hydrocarbons in commercial fuel gases that two trillion cubic feet per year of methane, ethane, propane and butane are available for conversion into oxidation products. It has been estimated that, if this oxidation could have been carried through to the production of alcohols, it would have resulted in the theoretical production of some 30 billion gallons of methyl, ethyl, propyl and butyl alcohols.[6] If, in 1929, the importance of these reactions was seriously being considered, the situation today is three or four times as great. The natural gas reserves in the United States have now increased to the point that they are approximately 160 trillion cubic feet, almost 400 percent more than the amount available in 1930. While the value of butane and propane has increased to the point where these gases are no longer considered low cost raw materials for oxidation, natural gas still represents the greatest raw material reserve for the synthesis of oxygenated compounds.

Possibilities Not New

The theoretical possibilities of oxidizing hydrocarbons to valuable products have probably existed since the chemical structure of natural gases was understood. One of the most important milestones in the development of direct oxidation of hydrocarbons stems from the patents of Glock in 1898-1899.[4] From that time to the present time, it has been claimed that many compounds, such as ethylene oxide, acetaldehyde, formaldehyde, propionaldehyde, propanol, butanol, ethanol, methanol, methy ethyl ketone, acetone, acetic acid, formic acid and many others, have been made by the oxidation of hydrocarbons.

The mild oxidation of aliphatic hydrocarbons yields, as principal products, methanol, formaldehyde, acetaldehyde and acetone, as well as smaller quantities of acids, higher alcohols and aldehydes. The amounts of each of these oxygenated products vary as to the type of hydrocarbon used, as well as the operating conditions employed for oxidation. Table 1 lists the principal oxygenated products of the lower aliphatic hydrocarbons in their order of decreasing yields.

TABLE 1

Hydrocarbon	Oxygenated Products* In Order of Decreasing Yields
C_1	2, 3
C_2	3, 2, 1
C_3	1, 3, 2, 4
n — C_4	1, 3, 2, 4
i — C_4	4, 1, 3, 2

* Key to Table No. 1

1—Acetaldehyde
2—Methanol
3—Formaldehyde
4—Acetone

When it is considered that the vapor phase oxidation of hydrocarbons can (at least theoretically) produce any of the various intermediates before reaching the ultimate end products, carbon dioxide and water, it is not surprising that various investigators have claimed the production of one or more of these chemicals from hydrocarbon oxidation. However, instead of many possible products being a stimulant to the commercial production of oxygenated compounds, it has actually resulted in the hindering of the commercial progress. What we might term "incidental" exploration of the reaction has too often led to interpretations completely foreign to the best practices necessarily employed for producing desired end products. This is probably best illustrated by the evident analogy in the procedures employed. For example, the usual means of oxidizing methane is by using a so-called oxidation catalyst, the most commonly employed being copper or copper-bearing compounds. While copper may be a fairly good catalyst for effecting the oxidation of methanol to formaldehyde, the assumption that it will be superior for methane oxidation is not based on a clear understanding of the kinetics of the reaction nor the stability of the products resulting from oxidation.

In not all instances has the study of catalysts been

the only source of misinterpretation. In many instances, investigators have elaborated upon results derived through long periods of heating and exposing reactants and products to severe operating conditions. In terms of industrial oxidation these results mean nothing. It is apparent that such approaches could lead only to low yields of desired products and would promote the formation of end products, carbon oxides and water. Other products obtained by such means would be characterized by extreme stability or would be equilibrium products. The most desired products obtained commercially are not characterized by either of the above properties.

First Commercial Unit

Probably, as a result of the above mentioned reasoning, the commercial aspects of the direct oxidation of the lower aliphatic hydrocarbons have languished. In the years from 1898 to about 1925 there seems to be little to indicate much progress from the commercial point of view. Actually, in 1925, only 1/10 of one percent of the synthetic aliphatic organic chemicals produced in the United States were derived from petroleum. Investigators and their results seem to have played only a small part in the development of the first commercial hydrocarbon oxidation unit. This was built by Cities Service Oil Company at Tallant, Okla., in 1926 and was an outgrowth of attempts to render corrosive oxygen-bearing natural gases noncorrosive. This plant was an important step forward in the commercial oxidation of hydrocarbons.[1, 8, 9]

Cities Service Oil Company has been the pioneer in the direct oxidation of the lower paraffin hydrocarbons to valuable oxygenated chemicals. J. C. Walker, one of the authors of this paper, and his associates discovered in 1925 that the corrosion of pipe lines caused by gas containing free oxygen could be minimized by bringing the gas to the necessary temperature and pressure to permit the oxygen present to be completely used up by reacting with the hydrocarbons. The formation of liquid oxygenated chemicals accompanied this reaction. The first commercial plant using an oxidation process of this type was constructed at Tallant, Oklahoma, in 1926. This plant has been expanded and three additional oxidation plants have been built.[7]

Direct partial oxidation of hydrocarbons can be carried out with or without a catalyst by heating air in admixture with an excess of hydrocarbons to a temperature where the oxygen is consumed.

The operation of the Celanese Corporation oxidation plant at Bishop, Texas, as can be seen from the patent literature, makes a great forward step in control of the oxidation reaction by using various agents for this purpose. Some of these agents for controlling reactions are diluting steam, nitrogen and other such inert gaseous materials.

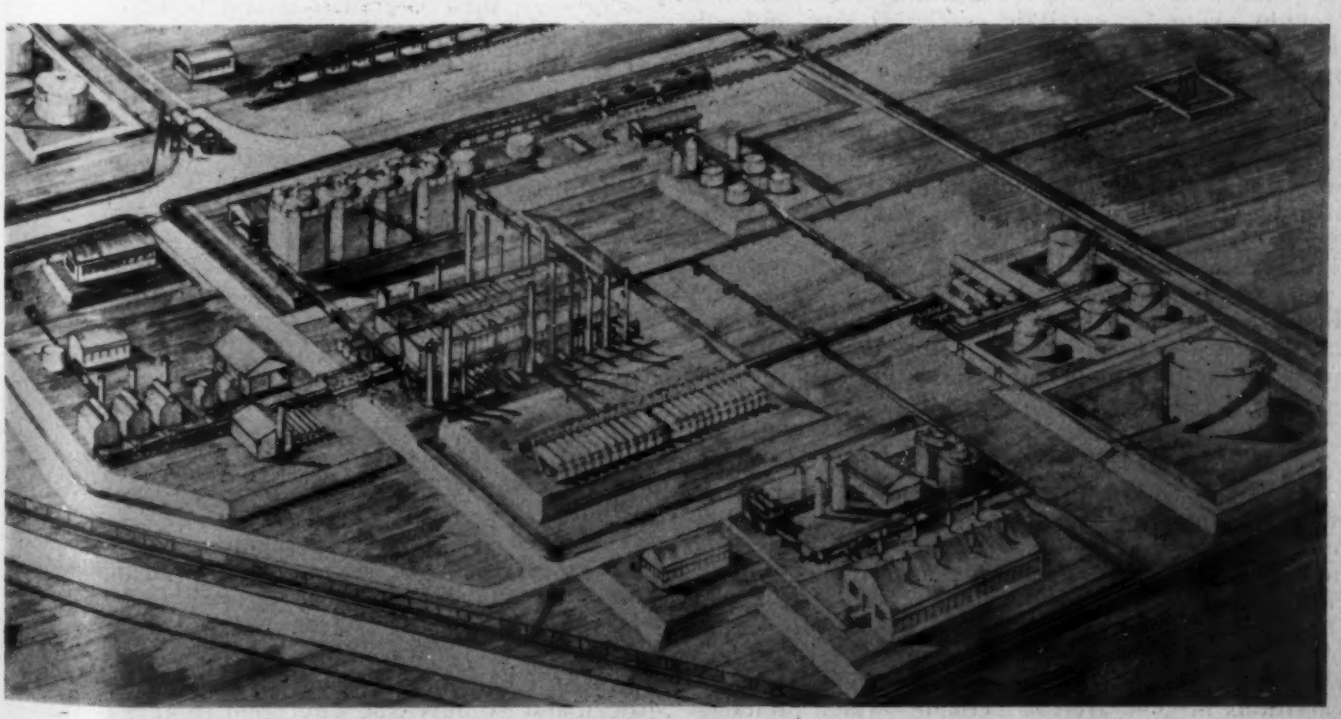
While the claims of the Cities Service Company's patents deal with the oxidation of methane, ethane, propane and butane, Celanese Corporation's claims are largely based on higher hydrocarbons than methane.

The importance of this industry can best be illustrated by the fact that Celanese Corporation at the present time is one of the largest producers of methanol and formaldehyde as reference to newspaper statements will bear out. Also, it might be pointed out that in 1945 the production of organic aliphatic chemicals from petroleum has increased to 50 percent, whereas in 1925 it was only one-tenth of one percent.

Array of Catalysts

Numerous patents have been issued claiming yields of oxygenated chemicals from hydrocarbons by the above procedure in some modified form. Most of the processes are very similar, yet employing various hydrocarbons in admixture with air which are generally passed over a catalyst bed. Among the catalysts which have been used are iron, nickel, copper, silver, gold, platinum, palladium, vanadium, chromium, molybdenum, manganese, their oxides, mixtures of their oxides, mixtures of the oxides and metals, aluminum sulfate and alkyl ethers. This list is not intended to be complete, but only to show that quite a formidable array of catalysts have been suggested.

From the experimental results obtained, it would seem that the first investigators of the oxidation of hydrocarbons were probably more familiar with the normal procedure for petroleum processing for the



Engineer's Drawing of the $15 million McCarthy Chemical Company Plant Now Under Construction at Winnie, Texas.

Cities Service Oil Company's Tallant, Okla., Chemical Plant. (Photo by Aerial Photo Service, Tulsa)

production of gasoline than for the production of heat sensitive organic compounds. As might be expected, some of the same techniques were used in the oxidation that were used in the art of cracking. Up until recent years the art of cracking depended primarily upon sufficient control of time and temperature in order that a product might be derived which would fall in a predetermined wide boiling range. It is quite true that the cracking processes, so developed, produced this broad boiling range product with the least amount of loss to fixed gases and carbon. Actually, it has not been until recently that extremely close control of the cracking reaction has been required. This exacting attention to heat rise at the point of instability of the cracking feed stock was brought about by catalytic cracking. Any unusual temperature rises would result in short cycles of operation and increased cost of processing.

With the latest processes for catalytic cracking for the manufacture of gasoline and allied products, temperature control has become of even more importance, and the units are so designed that at all times a most rigid control is exercised over the temperature in the catalyst beds. If the first hydrocarbon oxidation had followed the latest technique in petroleum cracking, in all probability its development as an industry would have been far more rapid, since the value of close control of reaction temperature is many times more important in oxidation than it could possibly be in the cracking operation.

Close Temperature Control Needed

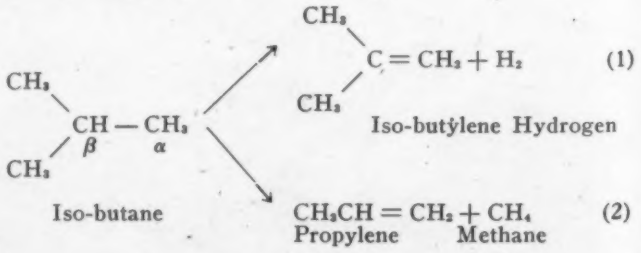
While it may be none too difficult to appreciate, we would like to illustrate, by use of the reaction mechanism and its close analogy to pyrolysis, the necessity of close temperature control during hydrocarbon oxidation.

It can be remembered that the yield distribution of various products shown in Table 1 vary widely with the hydrocarbon employed. The most marked variations in yield are for methane (which produces only methanol and formaldehyde) and iso-butanes. The iso-butanes yield substantial amounts of acetone in contrast to the other hydrocarbons. In fact, the whole degree of yields is different from those of the other hydrocarbons.

These variations in yield can be explained on the basis of the predominant bond scissions of the hydrocarbon employed. To arrive at the predominant bond scissions encountered in oxidation, the pyrolysis data of the hydrocarbon can be studied.

Regardless of the reaction mechanism involved, or the hydrocarbon radical in question, it is reasonable to believe that the degradation of hydrocarbon molecules and the position at which the reacting radicals attack the molecules is governed by the same factors that govern their degradation and point of attack during thermal pyrolysis.

For example, the pyrolysis of iso-butane in the lower range of pyrolytic temperatures yields propylene, methane and iso-butylene according to the following reactions:

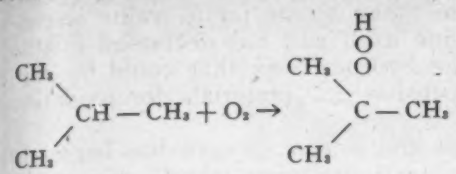


These two reactions amount to 85 to 95 percent of the reactions during low temperature pyrolysis of iso-butane.

The bond scissions predominating in pyrolytic cracking represent the weakest bonds in the molecule. These same bonds respond most readily and in corresponding degrees to oxygen.

In reaction (1) two carbon-hydrogen linkages are ruptured in the alpha and beta positions to give hydrogen and iso-butylene. This bond rupture is analogous to the oxidation mechanism where the carbon-

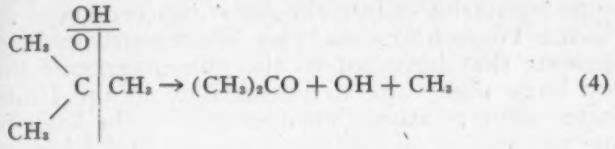
hydrogen group on the beta carbon is ruptured to form a "peroxide" as shown in the equation (3).



(3)

Iso-butane — Oxygen — Tertiary butyl hydroperoxide

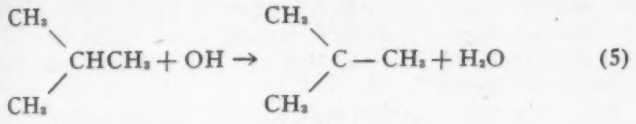
The "peroxide" shown in equation (3) disintegrates to a hydroxy radical (which attacks another iso-butane molecule starting a chain reaction), a methyl radical and acetone. The following series of reactions illustrate this:



(4)

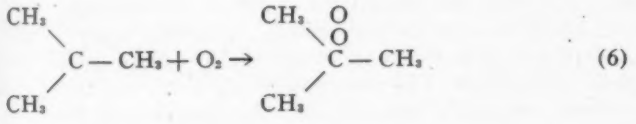
Tertiary butyl hydroperoxide — Acetone

The chain reaction is continued as follows:

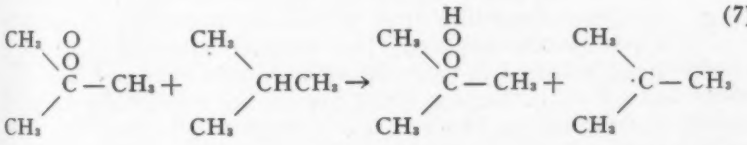


(5)

Iso-butane — Tertiary butyl radical

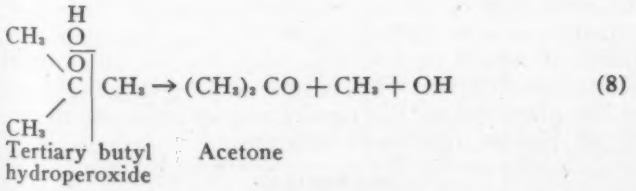


(6)



(7)

Iso-butane — Tertiary butyl Hydroperoxide — Tertiary butyl radical → O_2 → another chain



(8)

Tertiary butyl hydroperoxide — Acetone

The methyl radicals (CH_3) produced in equations (4) and (8) can be oxidized to produce formaldehyde:

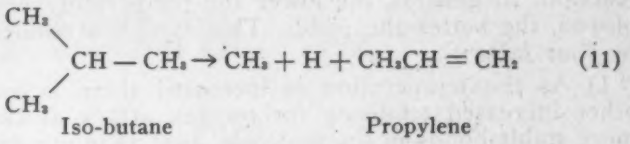
$$CH_3 + O_2 \rightarrow \underset{\text{Methylene Hydroperoxide}}{CH_2OOH} \rightarrow \underset{\text{Formaldehyde}}{CH_2O} + OH \qquad (9)$$

or combine to form, as in pyrolysis, ethane

$$2CH_3 \rightarrow \underset{\text{Ethane}}{C_2H_6} \qquad (10)$$

Going back to equation (3) it has been shown that the end products of the chain reaction which started with the carbon-hydrogen bond scission produced acetone and formaldehyde. This same carbon-hydrogen scission was also stated to occur in the pyrolysis if iso-butane in forming iso-butylene, substantiating the analogy between the predominant bond scissions during oxidation and pyrolysis.

Equation (2) requires the breakage of carbon-carbon linkage between the beta carbon and one of the methyl groups and an alpha carbon-hydrogen linkage. Equation (2) can be rewritten:



(11)

Iso-butane — Propylene

The methyl radical here, too, can form ethane or formaldehyde as in equation (9) and (10). The hydrogen can react in one of several ways—three of which are shown in equation (12-a), (12-b) and (12-c).

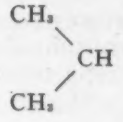
$$2H \rightarrow H_2 \qquad (12a)$$

$$H + O_2 \rightarrow OH + O \qquad (12b)$$

$$CH_3 + H \rightarrow CH_4 \qquad (12c)$$

and either become inert or furnish hydroxy radicals for the reaction.

If an alpha carbon-hydrogen linkage is attacked by oxygen, it can be shown by similar equations that the end products are again acetone and formaldehyde. The formation of acetone in the scheme involving the rupture of the alpha carbon-hydrogen linkage depends on the intermediate radical of the reaction



being attacked by oxygen to form a peroxide followed by degradation. However, this radical can alternately rearrange to form propylene. A series of similar reactions employing propylene produce ketene as a final peroxide degradation product together with a methyl radical and a hydroxyl radical. The ketene, being unstable in the presence of water and under the reaction condtions, yields with water, acetic acid; by thermal decomposition carbon monoxide and methylene radicals; and by polymerization diketene and other polymers. Diketene, if formed, would probably react with water to form carbon dioxide and acetone through the unstable acetoacetic acid. The predominating reaction would, however, be the hydration of ketene to acetic acid. This theory is a heretofore unmentioned mechanism for acid formation during hydrocarbon oxidation and is to a great extent substantiated by unpublished data which shows that acetic acid is the predominant acid formed during hydrocarbon oxidation, even though large

amounts of formaldehyde are always produced simultaneously.

Other series of reactions may be written which show the action of the hydroxyl radicals on propylene formed in reaction (2) to form acetaldehyde and methanol.

Lower the Temperature, Better the Yield

The most detrimental factor to good yields of these products is the lack of temperature control over the reaction. In general, the lower the temperature employed, the better the yields. This can be explained by four facts:

1) As the temperature is increased there is another increased tendency for oxygen attack at the more stable bonds in the molecule. Just as in pyrolysis, where the molecule has a greater tendency to split up (eventually to carbon and hydrogen) as the temperature is increased, the degradation of the molecule due to oxygen attack increases with temperature. The result of this increased degradation leads to products of lower molecular weight and increased quantities of fixed gases such as methane and the carbon oxides.

2) Methyl radicals tend to be oxidized too far due to hydroxyl radical attack.

$$CH_3 + OH \rightarrow CH_2 + H_2O \qquad (13a)$$

$$CH_2 + OH \rightarrow CH + H_2O \qquad (13b)$$

$$CH + OH \rightarrow CO + H_2 \qquad (13c)$$

This type of degrative oxidation of methyl radicals (or other radicals) is greatly favored by increased temperatures.

3) Increased temperatures increase the stability of the intermediate peroxy compounds with the result that there is an increased tendency toward "complete" decomposition.

$$CH_3 + O_2 \rightarrow CH_2OOH \rightarrow CO + H_2 + OH \qquad (14)$$

4) Fragile aldehydes, produced, as an end product of a chain reaction, tend to decompose to carbon monoxide, hydrogen and paraffins. This aldehyde decomposition is favored by temperature increases. There is no doubt that acetaldehyde and formaldehyde are extremely sensitive compounds and that their tendency to decompose to carbon monoxide and hydrogen or methane is greatly accelerated by temperature increases.

Thus, it becomes evident that the proper reaction temperature must be employed and precise control exercised over it.

Change in Status of Gases

An interesting economic situation has developed in connection with iso-butane. A few years ago isobutane, along with butane and propane, was considered a waste product in the petroleum industry. Some small portions of these products were disposed of by tank car shipments, but the larger portion produced in the gasoline plants was disposed of as fuel for the plant boilers and furnaces. In very recent years the consumption of these liquefied petroleum gases has increased to the point that they are no longer consdiered as waste products; to the contrary, they are now considered as valuable assets. This has necessarily placed the liquefied petroleum gases almost in the same category, as far as value is concerned, as gasoline itself and has decreased by the same number the hydrocarbons that could be considered as inexpensive raw materials for oxidation processes.

In recent years the role of oxygen has begun to assume even greater importance largely due to the developments during the last war of a cheaper procedure for the manufacture of relatively pure oxygen. By improving upon the methods used by Linde-Frankl in Germany, it has been stated that oxygen could be produced in the United States for a price as low as $3.50 per ton.[5] Immediate advantage has been taken of this rapid change in values of oxygen by incorporating it into the American version of the Fischer-Tropsch process. This has resulted in developments that have led to the announcements that two large plants are to be installed in the United States using relatively pure oxygen as the basis for the preparation of carbon monoxide and hydrogen for the process. It has been stated that, due to the advantages derived from the cheaper oxygen and the engineering improvements over the German methods, the plants can be built in the United States at the present time for approximately twice the cost per barrel that a refinery would cost. This is, of course, based on the consumption of natural gas as the raw material.[2]

The role that oxygen has played in this development is probably of greater significance than any of the engineering improvements on the Fischer-Tropsch process itself. In all probability, in the coming years the tonnage of oxygen so consumed will over-shadow any other single reagent now being used in the transformation of hydrocarbons into organic chemicals. The success of all this, we can rest assured, depends greatly on the degree of control that is exercised on the reactions involved. It is also our understanding that, while the presently considered modifications of the Fischer-Tropsch process have resulted in a much lower capital cost per barrel of oil so produced, there are in the formative stage developments being carried out in the United States at the present time that will produce similar results at both lower capital and operating costs.

While these developments may result in a tremendously increased value of gas properties to even the relatively small producer, to the technical man the point of greatest significance is that the American industry is now facing a new field of endeavor, the limits of which can be exceeded by the most optimistic imagination. This future will be largely based on the great steps that can be expected in the future of all phases of direct oxidation.

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Practical Considerations in Process Design of Fractionators

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IN THE process design of fractionators the engineer must use data obtained from a number of different sources. Proper use of this information and careful analysis of each feature of the design from a practical viewpoint are necessary in order to make a successful installation. There are numerous places to which the engineer may turn to find methods of solution for individual problems in fractionator design. However, there is no one source which he can use as a guide to point out the various problems he is likely to encounter. This article is intended to be such a guide. The practical considerations in the process design of large-scale fractionators are pointed out and discussed.

FRACTIONATION has gradually assumed more and more importance in industry until large fractionating towers are now familiar sights in petroleum refining and chemical plants. The use of such towers was augmented to a great extent during the recent war by the construction of aviation gasoline, styrene, and butadiene plants which require large numbers of fractionators in order to obtain satisfactory products. It is also expected that substantial numbers of fractionators will continue to be required by the petroleum and chemical industries.

Because of the large investment in fractionators and the importance of the jobs they must perform, considerable effort is expended in obtaining reliable designs. The responsibility of designing these large bubble-plate fractionating towers belongs to the process engineer. He usually can obtain data for his design from the laboratory, pilot plant, previous full-scale operations, or theoretical calculations. However, he must carefully weigh the data from each source in order to emerge with the final design which is most likely to result in a successful installation. Although theoretical calculations are quite valuable, the practical aspects in the design of a fractionator cannot be over-emphasized. The purpose of this discussion is to point out the design features which should be carefully analyzed from a practical angle. It is intended that it serve as a guide in fractionator design rather than furnish a means of solution of the problems encountered.

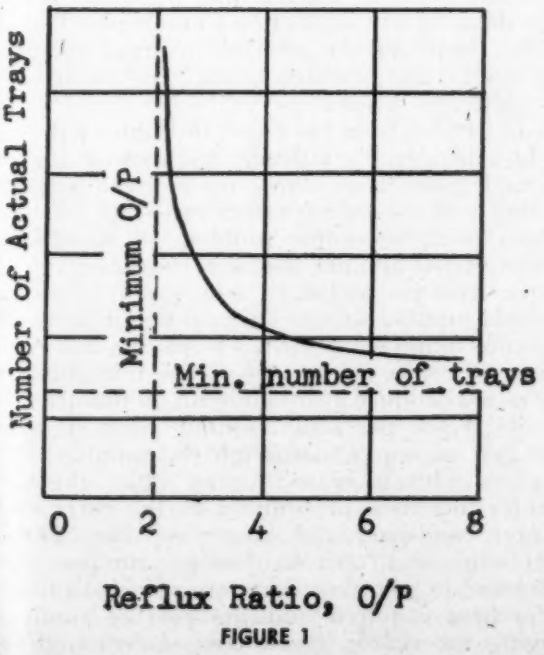

FIGURE 1

General Considerations

The total investment in even a small fractionator and its auxiliary equipment amounts to a sizeable sum of money, and so the engineer strives to design in such a way that the initial investment will not be too large. However, there are two other items which should sometimes be given even more consideration than the original investment. These are operation and maintenance. Considerable flexibility for these two items must be allowed for in the design or their costs may easily exceed the investment cost in a short time. Inability of a tower to make a satisfactory separation of to handle the volume of material processed by the remainder of the plant can result in large production losses. Loss of production due to down time, and large repair costs will result if an undue amount of maintenance is required. In addition to the above losses, a sizeable error in the original design may eventually necessitate shutting down part or all of the plant in order to make revisions or additions.

The process engineer is usually fortunate enough to have a number of places to turn for information upon which to base his design. Needless to say, none of these places should be overlooked. In addition to literature in books and periodicals, the engineer can normally obtain considerable data from the laboratory or research divisions of his own company. Quite often data from these sources, if properly interpreted, will lead to valuable operating and maintenance provisions which otherwise might be omitted. If improperly interpreted, data from the laboratory may later

result in many troubles which could have been avoided. If the process is a new one, a pilot plant may be constructed in order to obtain data more comparable to full-scale operation. Even here it must be remembered that troubles in the pilot plant will be magnified in full-scale operation and still other problems not encountered in the pilot plant may later cause troubles in the final plant.

One source of valuable information not always utilized to the fullest extent is the previous operation of similar units. In addition to consulting those men who engineered the units, the process engineer should also seek advice from men who have operated them and men who have maintained them. These men can usually give some valuable pointers to the process engineer because he does not always think along the same lines that they do. The mechanical design engineers can also aid the process engineer in numerous ways because they too look at the problem from a different angle. Safety features have quite often been neglected but are receiving somewhat more emphasis in recent years. It would be best to ask the safety engineer to check the design for unsafe practices instead of waiting for him to condemn the installation after completion. The process engineer himself should strive to acquire a "safety first" attitude in his work.

Although it should be the process engineer's job to design the unit, he should be eager to obtain information from the individuals and departments mentioned above. If he analyzes this information and incorporates it properly in the final design of the fractionator, the result will be a unit most satisfactory to all departments concerned as well as one which he believes will do the designated job properly. Also, inter-departmental strife will be lessened because all departments will feel that their ideas have received some recognition. The fractionator will be designed from both theoretical and practical considerations.

Salient Features in Column Design

Since the fractionating column is the principal piece of equipment in a fractionation unit, the salient features in its design will be the ones discussed here. Even though important, insufficient space is available to treat both the auxiliaries and the fractionator itself in one paper.

The general topics under which most column design features may be classified are, 1) material of construction, 2) operating pressure and temperature, 3) number of trays, 4) tray spacing and column diameter, 5) column height, 6) bubble tray, and 7) column openings. An analysis of the various items under these classifications will help to point out the importance of the practical aspect of the process engineer's work.

1) *Material of Construction*

Although it may not be the direct responsibility of the process engineer to specify the material of construction for the fractionator, he should concern himself with this problem from a process standpoint. Much of the data required may be obtained from the laboratory or pilot plant. Even though the metallurgist can specify the correct material for contact with components in the process, he may overlook other factors. Certain metals which may satisfactorily withstand corrosion may themselves catalyze undesirable chemical reactions within the system. The process engineer should see that the specified materials of construction will undergo little or no change themselves and that they will not be harmful to the process.

2) *Operating Pressure and Temperature*

If the fractionator is to be installed in an existing plant, it should be designed to operate at a pressure satisfactory for the process and to fit into existing facilities as easily as possible. An engineer may decide a certain column will perform best if operated at a pressure requiring refrigeration. If, however, the plant in which the column is to be placed has no refrigeration facilities but has an abundance of cooling water available, the engineer should consider operation at a higher pressure. In analyzing the two proposed operating pressures, he will also have to weigh such considerations as type and size of equipment required, process efficiency, and effect of pressure on equilibrium relationships of components, and ease of operation and maintenance. Sometimes the heating facilities rather than cooling facilities may dictate the operating pressure. If the proposed operating temperature required a direct-fired heater but a good steam supply is already available, the engineer may be able to re-design the unit to operate at a lower pressure and thus utilize the existing steam facilities.

3) *Number of Trays*

Determining the number of trays required for a fractionator is one of the most important jobs facing the designer, particularly if he has no previous operation upon which to base the design. A fractionator which has too few trays to make the required separation may prevent the making of a final product of specification purity or may necessitate the entire plant or units of it to operate at higher rates without a corresponding increase in final product volume. On the other hand, a fractionator with too many trays represents a waste of money in the initial investment, and in large installations this waste can be considerable. The process engineer should try to emerge with a fractionator which will do the designated job satisfactorily with as low an overall production cost as possible.

The number of trays in a fractionator cannot be calculated independent of the reflux ratio, for these two factors together are the ones primarily responsible for defining the separation to be made. The effect of reflux ratio on the number of trays required is shown in Figure 1, which is the type of curve such plots generally assume.

It can be seen from the curve in Figure 1 that there may be considerable latitude in choosing the reflux ratio and number of plates for a given separation. The choice of the proper reflux ratio and the number of trays is an economic problem of balancing the operating costs against the fixed charges on the fractionator and its auxiliary equipment. Use of the minimum number of trays would result in a column of infinite diameter in order to handle the required amount of reflux while use of the minimum reflux ratio would produce a column with an infinite number of trays. Thus, the actual column must be between these two extremes. Although the number of trays decreases with increased reflux ratio, the column diameter increases in addition to the extra cost for a larger condenser and larger reboiler. Thus, the fixed charges as a function of reflux ratio pass through a minimum. Operating costs are principally the cost of the heat required and the cost of cooling and pumping the reflux. These costs increase almost in

direct proportion to the increase in reflux ratio. Total costs, the sum of the operating costs and the fixed charges, also pass through a minimum as the reflux ratio is increased. A typical plot of fixed charges, operating costs, and total costs, as a function of the reflux ratio is presented in Figure 2. The higher the operating costs are relative to the fixed charges, the closer the minimum in total costs moves toward the minimum reflux ratio.

One factor in determining the number of trays which is often not given enough consideration is the relationship of the column in question to the remainder of the plant. Sometimes a little extra caution in designing one column may pay dividends in more efficient performance throughout the plant. A good example of this is illustrated in Figure 3 which is a simplified flow diagram of a plant manufacturing butadiene from normal butane. The function of Column A is to separate butene-1 from n-butane. The butene-1 is taken overhead and transferred to the butene dehydrogenation step for production of butadiene. The n-butane remains in the kettle product and is processed further, eventually being recycled to the butane dehydrogenation step. If Column A has an insufficient number of trays and as a result must be operated so that abnormal quantities of n-butane appear in the overhead product, a number of undesirable conditions will be created throughout the remainder of the plant in addition to the loss of some n-butane. N-butane in the butene dehydrogenation feed stream will increase, thus reducing equipment capacity there, and subsequently in the butadiene purification step. The only way this high concentration of n-butane can be reduced is by recycling a stream from the butadiene purification step to the butene purification step. However, such a procedure further loads the equipment in this step and reduces its capacity for processing fresh feed. Thus, it can be seen that the process designer must thoroughly analyze the relationship of each fractionator to the entire plant before releasing his final specifications. This same type of analysis should also be applied when considering design features other than the number of trays.

4) *Tray Spacing and Column Diameter*

The distance between trays is generally determined by theoretical calculations, experimental work, or comparison with similar processes already in operation. Since entrainment of liquid in the vapor rising from tray to tray is one of the chief factors in setting the tray spacing, the column diameter must be considered at the same time. Usually the tray spacing is set at a reasonable figure and the diameter calculated rather than setting the diameter and then calculating the tray spacing. However, the height of the fractionator must also be considered. If a 30-inch tray spacing would require too high a column for one section and insufficient room were available for construction of another section, then perhaps the designer would want to use smaller tray spacing and compensate for it by increased diameter.

The relationship of the fractionator to the remainder of the plant should also be subject to close scrutiny when determining the column diameter. Even though a column makes a satisfactory separation, it may be a plant bottleneck if it has insufficient capacity. Each fractionator should be in balance with the rest of the process equipment so that the entire plant may handle approximately the same load. Too small a column may mean loss of production and too large a column, wasted investment.

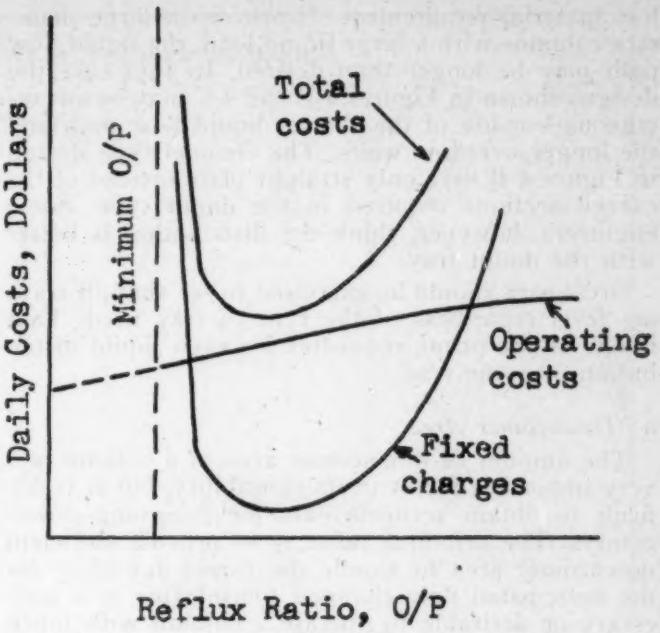

FIGURE 2

5) *Column Height*

After setting the number of trays, tray spacing, and column diameter, the designer generally has already determined the approximate height of the column. There are other features, however, which must be designed before the final height of the column can be set. If internal reboilers are to be used, the bottom part of the column must provide sufficient space for the reboilers and for disengagement of vapors below the first tray. If external reboilers are used, the column must have enough space to maintain the proper liquid head and to permit the installation of the return vapor nozzle from the reboiler. Additional liquid head is often obtained by placing the column proper on a high foundation.

In some processes, undesirable chemical reactions are promoted by keeping liquid at a high temperature in the reboiler. For example, in an extractive distillation process using furfural, the time and temperature at which the kettle liquid is kept has a definite effect on the amount of furfural polymerization. In such cases, it is well to design the reboiler so that the liquid hold-up time will be as short as possible.

6) *Bubble Tray*

The bubble tray in a fractionator is one of the most important features of design with respect to actual operation and maintenance. Yet, many significant points are often given little or no consideration in the original design. Too often the tendency has been to use blindly the same tray design already in use on other columns even though the new process is markedly different from the existing ones. Previous experience should certainly be drawn upon, but not to the exclusion of analyzing the design for possible improvements.

Skeleton sketches of a few types of trays are shown in Figure 4. A close analysis of each type would reveal certain advantages and disadvantages which each possesses. It is the job of the process engineer to weigh the various points and decide which tray he thinks will be best for his particular process. For example, the type shown in Figure 4-A has a

low material requirement. However, in large diameter columns with a large liquid load, the liquid flow path may be longer than desired. In that case the designs shown in Figures 4-B and 4-C may be advantageous because of the shorter liquid flow path and the longer overflow weirs. The channel type design in Figure 4-B uses only straight plate instead of the curved sections required in the donut type. Some engineers, however, think the distribution is better with the donut tray.

Great care should be exercised to see that all trays are level regardless of the type of tray used. That is one of the prime requisites for even liquid distribution over the tray.

a) *Downcomer Area*

The amount of downcomer area in a column is a very important factor in its operability, but it is difficult to obtain accurate data for designing downcomers. The principal point is to provide sufficient downcomer area to handle the liquid downflow for the anticipated flow changes. Quite often it is necessary or desirable to operate a column with much higher feed or reflux rates than contemplated in the original design. Previous operation of columns handling similar materials is the best guide. If such information is not available, the engineer must then rely either on pilot plant data or his own calculations. The characteristics of the liquid may often be overlooked, but are quite important. A liquid which tends to foam will require more downcomer area than one which does not, if flooding in the column is to be prevented.

b) *Weirs*

Most trays are designed with both inlet and outlet weirs. The length and height are the factors to consider in weir design. The height of the inlet weir should be just high enough to insure a positive liquid seal in the downcomer at all times. Too high a weir will create too much liquid hold-up in the downcomer and may necessitate larger tray spacing. Too high an outlet weir will increase the pressure drop across the tray unnecessarily. The height of this weir should be just enough to provide a positive liquid seal above the slots of the bubble caps. The length of the weirs may already be set by the downcomer area. However, some variation in weir lengths can be obtained by the choice of the type of tray as pointed out previously. For columns with very large liquid downflows, the channel tray offers approximately twice the weir length of the segmental type tray. When the liquid downflow is small, better distribution across the weirs may be obtained by notching. Great care should always be taken to insure that the weirs are straight and level to aid in obtaining proper liquid distribution across the tray.

c) *Bubble Caps*

The type bubble cap to use in a fractionator is perhaps the most controversial subject in the design of a column. Hundreds of different caps have been designed, each supposedly having the proper construction for increasing column efficiency or producing some other beneficial results. Unfortunately, however, the design of bubble caps is still far from exact.

A number of factors to be considered in selecting a bubble cap can be pointed out.

Since the bubble cap is in direct contact with both liquid and vapor (the latter often at high velocities), it may have to be constructed of alloy material in order to withstand corrosion or erosion even though the remainder of the column is carbon steel. This point is especially important in large installations where decreased tray efficiency as a result of cap corrosion could bring about appreciable production losses.

By determining the size, shape, and slot area of the bubble cap, the engineer can almost complete its design. Circular caps of approximately six inches diameter are in popular demand because they offer only small resistance to liquid flow and are a convenient size for obtaining good vapor distribution on the tray. However, as mentioned previously, there are a large number of different types of caps in successful use. Data on the relative efficiencies of the various types are practically non-existant or unavailable for publication. The slot area of the bubble cap is usually set by allowing a certain vapor velocity through the slots after having already decided upon the type and number of caps to use. In this connection it is well to consider the possibilities of the slots becoming plugged with chemical residues, polymer, or coke. Where such a situation might exist the slots should not be so narrow that they are easily plugged, thus reducing column capacity and necessitating frequent shutdowns. Removal of the obstructions in the slots is also made easier by use of a wider slot in such cases. It is also desirable to give special thought to the selection of slots in those cases where the liquid on the tray tends to foam. It might be difficult to arrive at a decision in that instance, but there are many engineers who feel that the type of slot and area have a definite effect on the foaming tendencies of the liquid.

In large-diameter fractionators, the liquid flow path across the tray may be quite long. There is some evidence that such a condition may create a large hydraulic gradient across the tray and result in uneven vapor distribution. This latter situation in turn will cause liquid dumping through the risers nearest the inlet weir. The problem has been at least partially solved in some cases by installing caps with progressively smaller slot area, in the various rows from the inlet weir to the outlet weir. At any rate it is a factor which the process engineer should consider before proceeding too far with the design.

After the type and size of cap have been set, the usual practice is to try to install as many caps on each tray as possible in order to obtain good vapor distribution. The main point to consider here is to allow sufficient space between the caps so that liquid flow will not be unduly retarded and so that installation and maintenance work on the caps will be easy.

The majority of bubble caps are presently being constructed of light gauge metal and consequently require some means of anchoring them to the tray. This is sometimes done by welding the cap to the tray surface or more frequently by means of a stem extending from the top of the cap down through the chimney where it is bolted or wedged in place or attached to a holddown bar to permit easy removal of the cap. It is particularly important to install removable caps in those instances where fouling is likely to occur. Otherwise, maintenance work for cleaning will be very tedious and expensive and there will be undue loss of valuable operating time. In view of the fact that unknown factors affecting polymer formation may exist in the process or the column may be used later for a different process, it

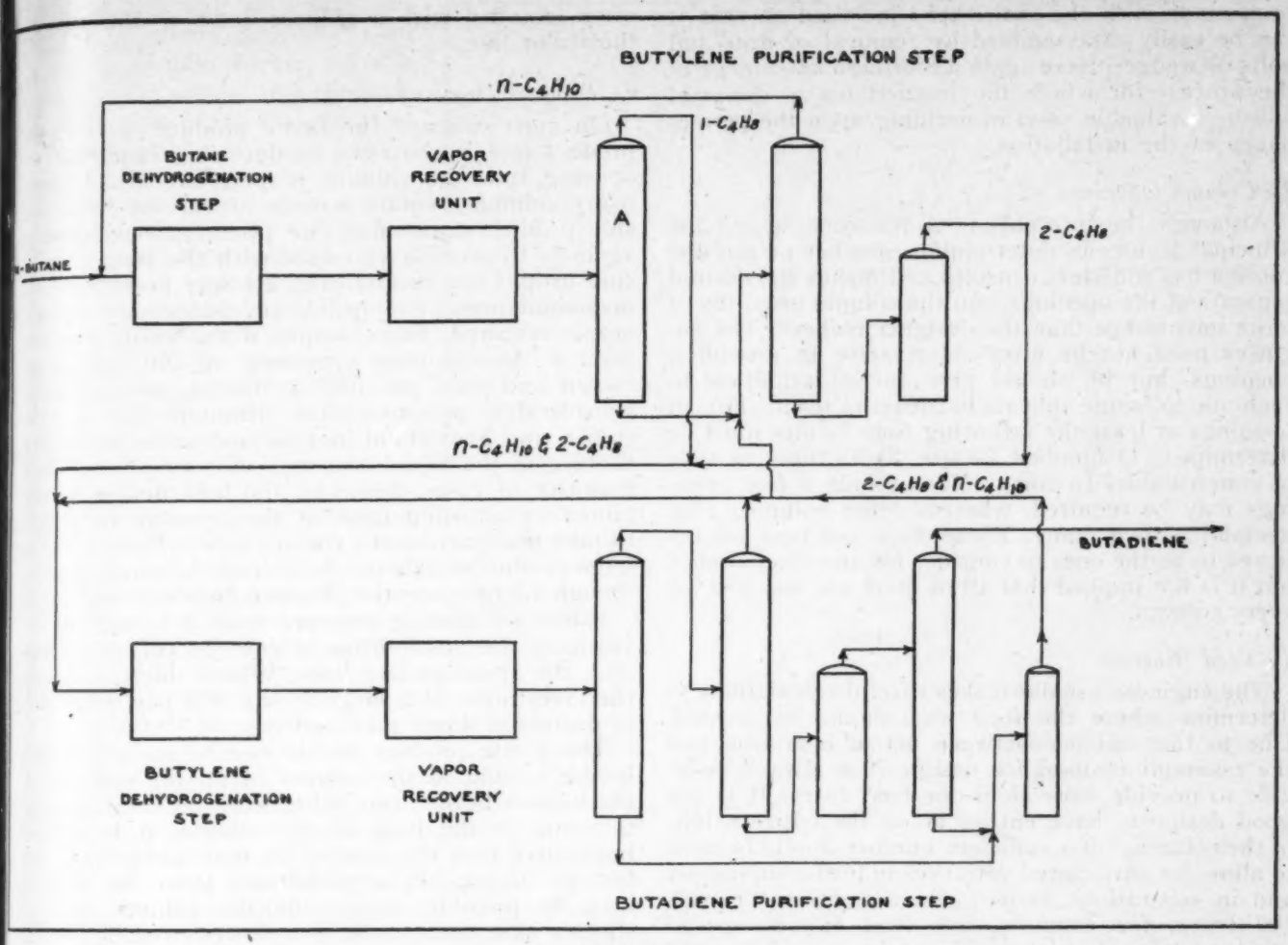

FIGURE 3

is always safest to install removable caps in any case. Exceptions to this might be in the case where a fractionator is designed for an identical process which has already been used successfully for an extended period of time.

It must also be decided whether to install the cap with the slots resting on the tray, or with the slots a definite distance above the tray. The only real basis for determining this is previous operating experience or pilot plant test work. In some cases, ease of cleaning polymer from the caps and trays may be a deciding factor in setting the cap with the bottom of the slots above the tray.

d) *Chimneys*

There are four decisions to make in designing the chimneys or risers: material of construction, diameter, height above the tray, and method of support. All of these are usually settled easily once the bubble cap design has been completed. Chimneys are generally made of the same material as the caps. The cross-sectional area of the chimney should be such that no undue pressure drop occurs through the chimney itself or through the reversal area between the chimney and the cap. The height of most chimneys is such that the top is just above the top of the slots on the caps. This dimension will be set once the size and position of the cap is determined. Chimneys, like caps, may be installed either in a removable fashion or rigidly attached to the tray. This latter method is carried out by welding or by rolling into the tray openings. The installation method used for the caps is the criterion for installing the chimneys.

e) *Tray Construction*

Much has been said so far about the attachments to the tray but no consideration has been given to the tray proper. Aside from the downcomer and riser openings, a well designed tray will have two other sets of openings. First, small weep holes should be provided so that liquid can drain from the tray when the column is down for internal maintenance work. These holes should provide sufficient area so that the tray drains easily and the holes are not easily plugged, but they should not allow enough liquid to drain through the tray to interfere with normal column operation. It is well to point out here that in special installations such as a water draw-off tray in a stripper the installation of any weep holes might be detrimental to operations. Second, each tray should be equipped with a removable section for use as a manway going from one tray to the next for repair work, unless a manhole is provided on the column shell at each tray. This may necessitate cutting out an odd shaped section from the tray or changing slightly the position of some of the bubble caps.

The diameter of the tray is a major consideration in determining its method of support. This question should be in the hands of the mechanical design engineers with the consultation of the maintenance section and process engineer. Common practice has been to either weld or bolt the tray in place. In recent years, however, a number of installations have

been made with the entire tray installed so that it can be easily disassembled by removal of nuts and bolts or wedges. Here again a thorough knowledge of the process for which the fractionator is designed will be a valuable asset in deciding upon the permanency of the installation.

7) *Column Openings*

Although the internals of a fractionator are the principal factors in determining whether or not the column has sufficient capacity and makes the desired separation, the openings into the column are often of more importance than the designer realizes. The engineer need not be ultra-conservative in providing openings, but he should give special attention to each one to insure that no bottlenecks result. For all openings at least the following four factors must be determined: 1) number, 2) size, 3) location, 4) type of construction. In some columns only a few openings may be required, whereas other columns may demand a great many. Those discussed here are believed to be the ones to consider for any fractionator, but it is not implied that all of them are required for every column.

a) *Feed Entries*

The engineer usually makes careful calculations to determine where the feed tray should be located. Due to the variance between actual operation and the assumptions used for design, it is always desirable to provide more than one feed entry. It is not good design to have entries along the entire height of the column, but a sufficient number should be used to allow for anticipated variances in feed composition and in separations desired. On a fractionator with 100 trays, for example, five feed entries spaced at four-tray intervals would not be out of line. In deciding upon the size of the feed entry, the engineer should check to see that undue pressure drop is avoided. Too small an entry could conceivably reduce capacity or at least require unnecessary power for pumping. Unlike some other openings in a fractionator, the feed nozzle cannot always end on the inside edge of the column shell. The feed nozzle may have to be extended well into the column in order to allow the feed to enter in such a way as not to disturb equilibrium on the trays. This will, of course, depend almost entirely on the types of downcomer and tray layout used. In this connection, it should be remembered also that during actual operation it may be necessary to introduce the feed either as a liquid or a vapor or mixture of both. Since the nozzles on a column may be subjected to more severe conditions than the column shell proper, it is often general practice to specify heavier material for the nozzles. This is true particularly of small nozzles such as sample nozzles, but might also apply to feed nozzles and others.

b) *Overhead Vapor*

The overhead vapor nozzle should be large enough so that the pressure drop through it will be quite small. Because of the wide variances in reflux volume which are often used, it may be desirable to specify this nozzle to handle a vapor volume considerably larger than the design volume.

On large columns the overhead vapor nozzle may serve also as a manhole for entering the column for maintenance work. This is accomplished by the installation of a hinged ell. Even when a relatively small overhead vapor line is used, a large nozzle is often installed with a reducer down to the size of the vapor line.

c) *Product Draw-offs*

On most columns the kettle product or bottoms product line is the only product line requiring an opening into the column proper, but in addition, many columns require nozzles at various trays for side product draw-offs. The primary consideration again is to provide an outlet with the proper pressure drop. Only in this case, a larger pressure drop may sometimes be allowable and hence only a small nozzle required. For example, if the kettle product from a depropanizer operating at 300 psi. were cooled and then put into a 100-psi. storage tank, considerable pressure drop through the nozzle, cooler, and lines could be tolerated unless excessive flashing of the liquid occurred prior to cooling. In a majority of cases, however, the best design is obtained by allowing most of the pressure reduction to take place across the control valve. Thus, a fairly large product nozzle on the column is required even though a large potential pressure drop is available.

When considering pressure drop it is well to investigate the possibilities of coke or polymer plugging the openings and lines. Where this can occur, the investment in a large nozzle will pay dividends in decreased down time and ease of cleaning.

The kettle product nozzle can be located either in the bottom of the column or on the side above the bottom head seam. When very heavy ends tend to settle in the base of the column, it is almost imperative that the product be withdrawn from the bottom. If product is withdrawn from the side, it may be possible to provide the column with a shorter skirt and result in a lower overall column height.

Nozzles for liquid side-product withdrawal should be installed in a downcomer and close enough to the bottom of the downcomer to insure that they will always be covered with liquid.

d) *Reflux*

Here, as in the case of the overhead vapor line, it may be desirable to provide a nozzle suitable for much larger volumes than the design volume. Pumping costs are a prime factor in handling reflux and therefore no great restrictions to flow should be installed in this system.

The reflux should be introduced into the column in such a manner that even liquid distribution is obtained on the top tray. Unequal distribution on this tray could upset the equilibrium on a number of trays below.

e) *Relief Valves*

Even though relief valves are very expensive the process engineer should adopt a conservative attitude in making provisions for them. The nozzle should be large enough for the standard size valve selected and in some cases an extra nozzle is provided for the possible addition of a second valve later.

f) *Manholes*

On large diameter columns manholes must be provided for entry into the column to do maintenance work. If each tray is equipped with internal manways so that a man can go from one tray to the next inside the column, the manholes opening into the column need be provided only at intervals up the

column. Intervals of either five or ten trays are generally used for they enable workers to distribute themselves well throughout the inside of the column when a large number of trays require work or afford a conveniently sized section when only a few trays need cleaning or repair. Where a column is in a service conducive to extensive coke or polymer formation, it may be good practice to provide manholes at intervals of less than ten trays. If there is believed to be little or no possibility of tray fouling or of corrosion necessitating frequent maintenance work, only a few manholes are required for a tall column. In no case, however, is it considered wise to install less than two manholes per column, one at the top and one at the base.

Manholes represent considerable investment and so should be kept as small as practical but as the name implies they must be large enough to accommodate a man safely. The size of a manhole is standard in most plants, but in case any question arises about the size consult the maintenance department for it is their men who must use the manholes.

The blind flanges on manholes are very heavy and therefore must be equipped with hinges so the maintenance men can easily swing them from the opening. It is highly important that these flanges be properly sealed when closed to prevent leakage which might result in a fire or explosion or at least a loss of product. Again the process engineer should consult the maintenance department for advice.

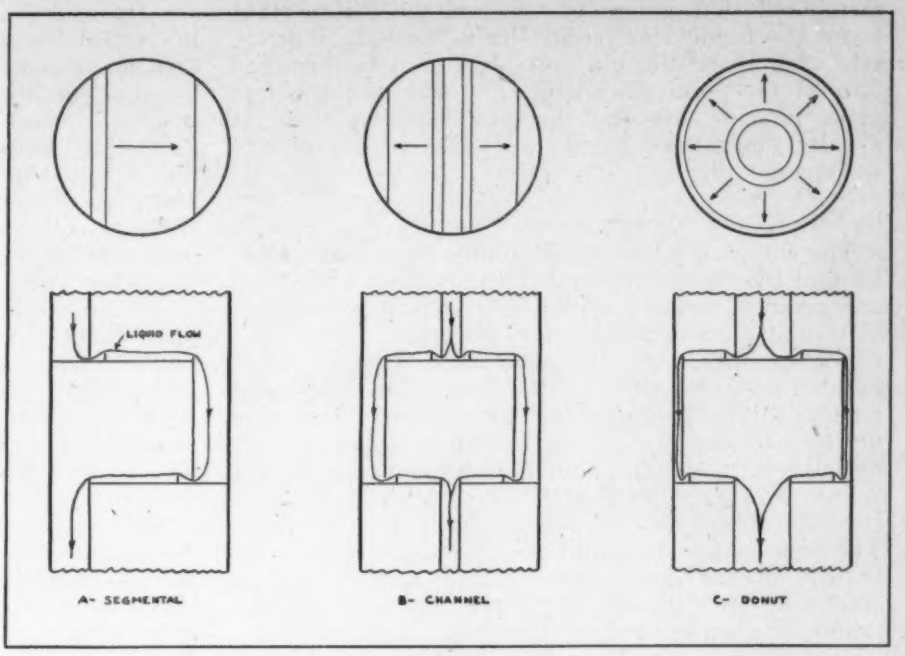

FIGURE 4

g) *Internal Heating Elements*

A great number of fractionators are designed with internal heating elements. When this is done there are several important factors to be considered. Since heating elements may require rather large column openings, the mechanical design section should be consulted to determine the largest size opening allowable before proceeding too far. If two elements are required, their relative positions must be decided. Metal supports and rollers should be installed inside the column to facilitate easy installation and removal of bundles. The top of the heating element should be far enough below the lowest tray to allow a controllable liquid level to be maintained above the elements and to keep liquid entrainment in the vapor entering the first tray at a minimum. Too much conservatism in this regard, however, will result in wasted investment of additional column height.

h) *Lines to External Kettles or Exchangers*

When heat is supplied by external kettles or exchangers, liquid and vapor lines between them and the column must be provided. The main consideration here is to be certain that the lines are large enough and as straight as possible to preclude the possibility of excessive pressure drop, thus limiting column capacity. The possibilities of fouling of these lines should be given careful thought. The nozzle for the liquid line from the column should be low enough to take full advantage of the liquid head in the base of the column. The nozzle for the vapor entry into the column must be far enough below the first tray so that vapor will be distributed evenly as it enters the vapor risers on the tray.

i) *Observation Ports*

There have been several instances in recent years where glass observation ports have been installed on large hydrocarbon fractionating columns. At least two nozzles are usually required, one for light and one for observation. Observation ports are of considerable value for studying tray action to find ways of correcting faulty operation of columns already installed and to collect data for future design work. The principal factors in installing observation ports are to locate the nozzles in positions where the tray can be observed best, provide easy to operate and accessible valves, and furnish a good source of light. The size of the openings required depends primarily upon the column diameter.

j) *Liquid Level Control*

The type of nozzle to install for a liquid level controller depends upon whether an internal or external float is used. In either case the engineer should follow the recommendations given by the manufacturer of the controller unless previous experience has produced additional refinements. Where considerable turbulence is likely to occur, it is often desirable to install baffles around the nozzle. When internal heating elements are used, the use of an external leg on the side of the column may be advisable. Installation of the liquid level controller on this leg will result in much smoother performance. Care should be exercised in locating nozzles because most liquid level controllers provide only small flexibility in changing the point of control.

k) *Gauge Glasses*

Sufficient number of nozzles for gauge glasses must be supplied to give adequate observation of the anticipated liquid levels. This generally means provision of nozzles from near the bottom of the column to within a few feet of the first tray. It should be re-

membered that gauge glasses are most important when the liquid level controller is working improperly and, therefore, glasses should not be installed only at the point of control. The size of the gauge glass nozzles and the distance between them is usually determined by the selection of a standard size gauge glass.

l) *Sample and Pressure Connections*

The complete story of a column's operation cannot be told by the volume and composition of the feed and product streams alone. To obtain further information on column performance, samples of liquid and vapor, or both, must be taken from the trays. The number of such sample connections to provide is left wholly to the judgment of the engineer. It should not be necessary to provide openings on every tray of a large fractionator, but liquid and vapor sample connections at every fourth tray, for example, on a 100-tray column would certainly not be too many. The best spot for a liquid sample nozzle is near the bottom of the downcomer. Vapor sample nozzles should be high enough above the tray to insure the taking of a vapor sample under normal operation.

Sufficient nozzles for pressure connections should be provided to enable the taking of adequate pressure drop data. Since sample nozzles can also be utilized for taking pressures, the only specific pressure connections required may be at the top and base of the column. The one at the top can even be eliminated by installing it in the overhead vapor line instead of in the column proper.

Because of their small diameter and the hard usage received by taking of samples or putting on and taking off pressure gauges, all sample and pressure nozzles should be made of extra heavy material regardless of the column operating pressure. The designer should guard against using too small nipples to provide adequate strength, e.g., they should be at least ½ inch diameter.

m) *Temperature Connections*

Every fractionator must be equipped with means for obtaining internal temperatures. This is done either by temperature indicators, temperature recorders, stationary thermometers, or thermowells with removable thermometers. The selection of the number, type, and location of temperature connections should be done by the process engineer, but it would be wise to consult the operating department also since their men will use them most. In locating connections for indicating and recording points to be used for control, the engineer should give careful thought to the number required, for he may wish to change control points after operation has begun. Thermowells should be furnished on all trays which have sample connections so that complete correlation of operating data may be obtained if desired. Special care must be given to the location of temperature connections in the base of a column with internal heating elements to avoid the taking of erroneous temperatures.

8) *Miscellaneous Features*

These discussions have covered most of the important design features which can be classified under definite headings. In addition there are several miscellaneous items to consider which often are of considerable importance in fractionator designing.

One of these items is insulation. The primary concern of the process engineer regarding insulation for the fractionator is that of making certain it is of proper quality and thickness for keeping heat losses at a minimum. It is also used to prevent changing weather conditions from making the fractionator difficult to control. In some instances insulation is used as a safety measure to protect personnel. The insulation selected in any case should be as weatherproof and easy to maintain and repair as possible.

Every column must be equipped with ladders and balconies to provide easy access to the valves and manholes along the column. All valves required for normal operation should be as easily accessible as possible so that they can be reached quickly and safely in case of emergencies. Sample connections and thermowells should also be close to the ladders. Balconies should be located at every manhole. Provisions for these in the original design will make maintenance work easier and safer and eliminate the necessity of using temporary platforms when internal column repairs are required. It is well to consult the maintenance department regarding other special features which might be included such as hoists or A-frames for hauling material up the column. The maintenance men will naturally wish to include as many time and labor saving devices as possible to aid them. Some of these may be eliminated because of prohibitive costs. Others may be provided for satisfactorily by temporary construction at the time the maintenance work is required. In those processes, however, where considerable fractionator maintenance work is anticipated, permanent installation of certain features strictly for the use of maintenance personnel may pay dividends in relatively short periods of time.

Conclusions

The process engineer should endeavor to use efficiently all information available to him before releasing final specifications on a fractionator. Representatives of the research, laboratory, mechanical design, operating, maintenance, and safety divisions should be consulted. Each part of the column should be analyzed thoroughly from a practical viewpoint. Particular attention must be given to those features which could easily impair operating efficiency or capacity if improperly designed. Failure to make allowances for certain maintenance problems may result in large maintenance expenditures and loss of production due to off-stream time for cleaning, repairs, or revisions. The relationship of the fractionator to the remainder of the plant is also an important factor to consider. Careful interpretation of data from various sources and close scrutiny of the design from a practical angle will produce an installation most likely to operate satisfactorily.

ACKNOWLEDGMENT

The author expresses his appreciation to Phillips Petroleum Company for granting permission to publish this paper. Special thanks are extended to R. G. Boatright, H. L. Hays, and D. S. Sherwin, for their helpful criticisms made during the course of preparation of the manuscript.

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Lab Speeds Up Research Work

A MICRO-LABORATORY — where scientists work with doll-sized instruments — is being utilized by Gulf Oil Corporation in speeding up progress of petroleum research. A chemical workshop in miniature has been set up and work is being done with quantities of material ten to hundreds-of-thousands of times smaller than used in ordinary research. These smaller samples can be analyzed more rapidly than normal amounts and, too, laboratory space is saved. Serious explosion dangers are also eliminated.

Shown on this page are five photographs of this laboratory.

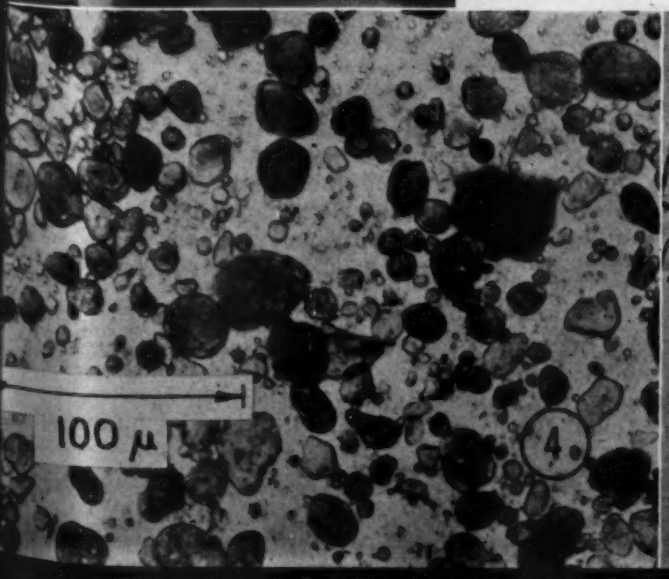
No. 1—A "boat" containing a sample of catalyst is charged into a miniature furnace. Purpose is to determine carbon, hydrogen and oxygen content, these being frequently needed in organic analyses.

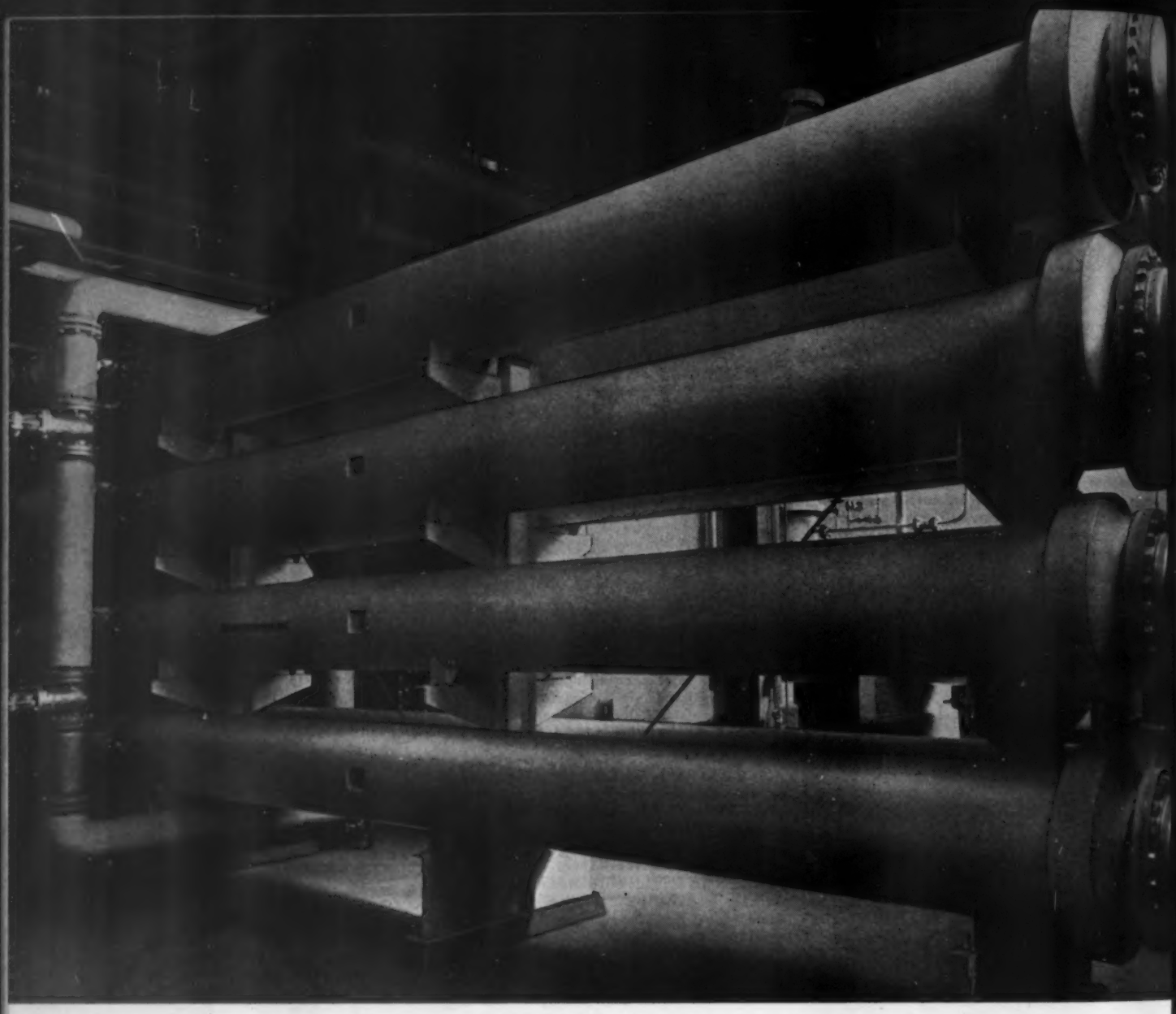
No. 2—A "micro manipulator." The hand lever operates a hydromatic drive which directs instruments so small they work within a drop of water as would ordinary tools in a beaker or test tube. Particles are moved about by infinitesmal probes, rakes, hoes and shovels. Work shown is being manipulated in a humidified cell, to prevent evaporation of the tiny sample as it lies beneath the microscope lenses.

No. 3—A "polaragraph" which can trace down one part in a million in a solution. Its analyses eliminate or shorten lengthy separation of materials and permit actual analyses in 10 minutes which otherwise might require as long as six hours. Operator watches instrument board as he handles controls regulating analysis.

No. 4—Photomicrograph of how a typical catalyst looks through the microscope lense. By studying size and other characteristics of catalysts at high magnification (shown is 500 times) technicians have been able to help isolate catalyst types which made modern refining more efficient. Note number of catalysts occupying space marked "100 u" (for microns) —the equivalent of 1/250th of an inch.

No. 5—A fraction of wax, these jagged particles being actually part of a commercial wax. They represent a crystalline wax fraction, magnified 500 times. Microscopic study of the forms of such materials often enables scientists to identify unknown particles by their structures.





Heat Exchangers in Alcohol Plant. (Photo courtesy Vulcan Copper & Supply Company)

Synthetic Alcohol from Petroleum

ROBERT S. ARIES, Consulting Engineer, Brooklyn
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INDUSTRIAL ethyl alcohol ranks high in the list of essential raw materials for the chemical process industries, both as a solvent and a starting point for many chemicals such as acetaldehyde, acetic acid, ethyl acetate, etc. Its methods of manufacture have thus attracted the attention of chemists and engineers for many years. Traditionally, alcohol is made by the fermentation of carbohydrates, either in the form of starches as in grain or as sugars in blackstrap molasses and sulphite waste liquors. However, the manufacture of synthetic ethanol from ethylene has become an increasingly important industrial process in recent years. In view of its favorable economics, the processes described as well as the production of by-product ethanol from the Fischer-Tropsch type synthesis using natural gas or eventually coal as a raw material, are slated to become increasingly important throughout the world.

Source and Purification of Ethylene

An average petroleum refinery employs mixtures of ethylene with other by-product gases as fuel for boilers. To make ethylene in quantity calls for adjustment in the usual refinery practices. This is im-

portant in the production of alcohol, because if the proper amount of ethylene is not produced as scheduled, the alcohol plant cannot operate on a definite time-table. Data on "once-through" fluid catalytic cracking operations may be interpreted to indicate that some 9 percent of the C_3—and lighter fraction consists of ethylene; the yield of this fraction ranges from 6.0-14.5 percent of the charge. Vapor phase cracking can yield a gas containing 20-25 percent ethylene, and higher percentages of ethylene can be obtained by cracking ethane-propane mixtures or the propane-butane cut of stabilizer gas. Fractions containing varying percentages of ethylene can be obtained in the course of operations designed for other products, such as butadiene and isoprene.

Provided that the ethylene content is sufficiently high, the saturated hydrocarbons, chiefly methane and ethane, need not be removed. However, most of the higher olefins must be removed to prevent excessive polymerization. In some cases it may be found desirable to crack crude petroleum, specifically for chemical by-products, thus getting a high percentage of purer ethylene which is well suited for conversion to alcohol.

Direct Hydration Process

The reaction $C_2H_4 + H_2O \rightleftarrows C_2H_5OH$ is exothermic, $-\Delta H = +10{,}964$ cals at 150° C. The values for the free-energy change, lead to the following values for log K: at 300° K, 1.119; at 400° K, —0.81; at 500° K, —2.02; at 600° K, —2.80. The formation of alcohol is favored by low temperature and high pressure, and the conversion of ethylene is favored by having a high ratio of ethylene to water, so far as it is economical to do this. At the lower temperatures the pressure cannot be increased indefinitely because the water would condense, thereby changing the equilibrium unfavorably, and at higher temperatures increase of pressure tends to cause polymerization of the ethylene. At 200° C., and with the water-to-ethylene ratio at 2 to 1, the equilibrium reaches 20 percent cenversion at 300 psi., but at 300° C., and with 5 to 1 water-to-ethylene, equilibrium conversion is only 15 percent at 1200 psi.

At the low temperatures necessary for favorable equilibrium the reaction rate is slow, and therefore catalysts must be used. Commercial utilization of the process has been proposed, presumably with efficient catalysts, although published data only indicate rather low conversions. At 290° C. and 1500 psi (conditions under which the equilibrium conversion is only 8 percent) conversions of 2-4 percent were obtained with a contact time of 60 seconds. British investigators have used, as catalysts, complex phosphates of many metals, including iron, nickel, calcium, strontium, barium and magnesium. Research in the United States found aluminum oxide the most active catalyst of about 20 materials, including oxides and phosphates of aluminum, thorium and tungsten.

Side reactions are not important at moderate temperatures and pressures. Some workers found polymerization at 350° C., but not at 320° C. or below, while others apparently did not get polymerization or ether formation at 150-300° C. and pressures up to 1500 psi.

Since the conversion per pass is necessarily low, recycling is required, and a comparatively concentrated source of ethylene must be used to prevent the building up of inerts.

Despite the somewhat adverse theoretical considerations such as the low conversion per pass, the direct hydration process presents some distinct advantages such as lower investment costs, no necessity for corrosion resistant equipment and the elimination of the acid solutions. Many of the world's leading chemical companies have investigated the process and Shell Oil Company is building a commercial installation utilizing this process, which will use substantially pure ethylene feed in order to recycle the unreacted ethylene. The operating costs of this process are expected to be similar to the ones using sulfuric acid.

Chemistry of Sulphuric Acid Process

The hydration of ethylene to alcohol can also be carried out indirectly, by first treating the ethylene with sulphuric acid, forming ethyl sulphuric acid (ethyl hydrogen sulphate) and diethyl sulphate.

$$C_2H_4 + H_2SO_4 \rightarrow C_2H_5HSO_4$$
$$C_2H_5HSO_4 + C_2H_4 \rightarrow (C_2H_5)_2SO_4$$

The ethyl sulphuric acid and diethyl sulphate, on treatment with water, hydrolize to ethyl alcohol and sulphuric acid.

$$C_2H_5HSO_4 + H_2O \rightarrow C_2H_5OH + H_2SO_4$$
$$(C_2H_5)_2SO_4 + 2H_2O \rightarrow 2C_2H_5OH + H_2SO_4$$

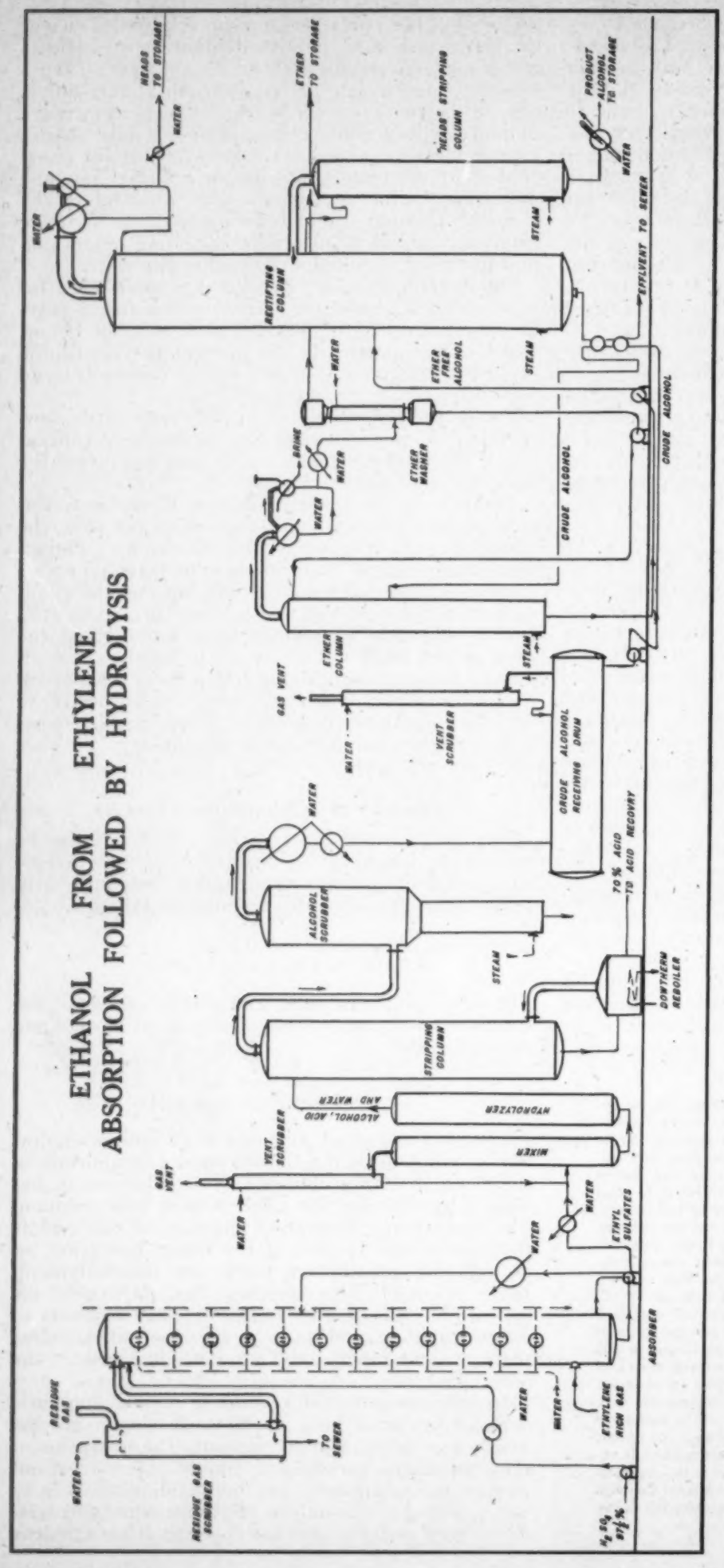
This process, which amounts to an indirect hydration of ethylene, is not limited by the equilibrium of the direct hydration, because the equilibrium is disturbed by diluting the concentrated acid solution. The free energy change of dilution of the acid is therefore added to that of the direct hydration, so that almost quantitative yields are thremodynamically possible. This process has, however, the undesirable feature that relatively large amounts of dilute sulphuric acid must be reconcentrated, unless there is some other outlet for the dilute acid in the same plant.

In order to avoid the reconcentration of sulphuric acid, efforts have been made to develop a process which uses dilute acid throughout. The formation of ethyl sulphates, and their hydrolysis, are carried out in the same equipment, and the sulphuric acid is, in effect, acting as a catalyst. High pressures and relatively high temperatures are required. Thus ethylene

ETHYLENE derived from petroleum cracking can be hydrated to give ethanol. Direct hydration is limited thermodynamically to rather low conversions. The reaction can be carried out indirectly by first treating the ethylene with sulphuric acid, thus forming ethyl sulphuric acid and diethyl sulphate, which, on treatment with water, hydrolyze to ethyl alcohol and sulphuric acid, which is concentrated and re-used. By-product ethyl ether is formed from the reaction between the ethanol and the unhydrolyzed ethyl hydrogen sulphate.

In this article the author stresses procedures now in commercial use in the United States. The reaction $C_2H_4 + H_2O \rightleftarrows C_2H_5OH$ is exothermic, $-\Delta H = +10{,}964$ cals. at 15° C. The equilibrium is favored by low temperature and high pressure. At 200° C., and with a water to ethylene ratio at 2 to 1, the equilibrium reaches 20 percent conversion at 300 p.s.i. Re-cycling is required, and a concentrated source of ethylene must be used to prevent the building up of inerts.

Commercial sulphuric acid and processes absorb the C_2H_4 in 90 percent acid or higher. The liquid product is hydrolyzed and stripped of alcohol and ether. After scrubbing with caustic soda, the alcohol is rectified, and a very pure product is obtained. Economics of the process are included in the presentation which originally was made to the International Congress of Pure and Applied Chemistry at its 1947 meeting in London.



and steam may be bubbled at 600-1000 psi., and 250° C. through sulphuric acid in a vessel packed with Raschig rings. The counter-current principle may be utilized, in which a large quantity of sulphuric acid is circulated, so as to keep the alcohol concentration low, thereby reducing ether formation. However, the high operating pressure necessitates special equipment, and very severe corrosion problems are encountered in handling 8 percent sulphuric acid at 250° C. This process is not operated commercially yet, although the relatively high conversions of 90 percent is claimed. The fact that impure gases can be processed and the elimination of the acid concentration step may in certain instances outweigh the disadvantages of the processes.

Concentrated Sulphric Acid Process

Concentrated acids are used by Enjay Company, Inc., of Standard Oil Company (New Jersey) and the several plants of Carbide and Carbon Chemicals Corporation.

This process may be compared with the similar synthesis of isopropyl alcohol from propylene, another refinery by-product. Propylene is, however, more readily absorbed in sulphuric acid; the more difficult absorption of ethylene necessitates special equipment to withstand corrosion. Many catalysts for this absorption have been tested and silver sulphate has apparently been found effective, as well as compounds of copper and of metals of the iron and platinum groups.

The ethylene-containing gases make contact with sulphuric acid, 90 percent strength or higher, in an absorbing tower. The tail gases from the tower, after scrubbing with caustic soda, can be used as fuel. The liquid product from the absorbing tower, containing ethyl hydrogen sulphate and diethyl sulphate, passes to a hydrolysis chamber, and then to a column in which it is further hydrolyzed and stripped of alcohol and ether vapor. The mixed vapors of alcohol, ether and water, after being scrubb with caustic soda, are condensed and sent to a crude alcohol storage tank. In the alcohol purification section an ether column takes off the ether, which is washed and sold as a by-product; the alcohol is then rectified in one column. A very pure product is made.

The sulphuric acid leaves the system as a black, dilute acid from the alcohol stripper. Reconcentra-

Condenser in Chemical Recovery Plant. (Photo courtesy Vulcan Copper & Supply Company)

tion of the dilute acid, in one American plant, takes place first in an acid reboiler, which brings the concentration up to 70 percent; the evaporated steam is returned to the stripper. For further concentration this plant uses a Simonson-Mantius two-stage vacuum system which concentrates sulphuric acid without causing objectionable fumes. In the first concentrator the acid is brought up to 89 percent under a 2.5 psi.; in the second it is brought to the final desired concentration at 0.4 psi. Because of the low pressures the acid temperatures are kept below 370° F., but since, even at this temperature, steam as a heating medium would require very considerable pressures, Dowtherm A vapor is used, which gives a temperature of 575° F. (thus providing ample temperature drop) at a pressure of only 20 psi.

The advantages of the process are the relatively high conversion to alcohol (70-80 percent) and the fact that gases containing as low as 30 percent ethylene can be successfully processed. On the other hand, it has some disadvantages, such as the concentration of considerable amounts of acid and the necessity of using special equipment. The equipment consists of steel pressure vessels with linings and internal parts of lead, ceramic materials, Karbate, glass, Hastelloy, Duriron, etc.

Flow Sheet of the Process

Following is a detailed description of a typical process for large scale production of ethyl alcohol from ethylene as illustrated in Figure 1. The process as shown, with certain variations in some cases, is representative of this process as practiced in the United States today. The normal yield of alcohol and ether from this process is more than 90 percent of that theoretically obtainable from the absorbed ethylene. Ether production amounts to more than 5 percent of the alcohol. Normally, the alcohol produced is of a denaturing grade; but, by modification of the distillation equipment, a higher grade of spirit may be produced.

Ethylene-rich gas, containing 32 to 100 percent ethylene, enters the bottom of the absorber and passes upward in a counter-current way to a descending stream of sulphuric acid and ethyl sulphates. The pressure in the absorber is 200 to 500 psi., depending

upon the concentration of ethylene in the feed stock. The absorber is operated at a temperature of approximately 180° F. Because the absorption gives off heat, it is necessary to provide for internal cooling coils to maintain this temperature. Liquid from the base of the column is also cooled and recirculated to an intermediate point in the column in order to maintain the desired temperature and ethyl sulphate concentration. Fresh sulphuric acid of approximately 97.5 percent strength (about 2 pounds for each pound of alcohol produced) is introduced into the top of the column. Residual gases from the top of the column are washed with water to remove entrained acid and pass out to other uses.

The liquor flowing from the base of the absorber is a mixture of diethyl sulphate, ethyl hydrogen sulphate and sulphuric acid. This liquid is cooled and passes to a mixing vessel in which an intimate mixture of this solution with water is obtained. The mixture then passes to a hydrolyzer in which hydrolysis is completed. The liquid leaving the hydrolizer consists of a mixture of ethyl alcohol, ethyl ether, sulphuric acid and water, together with a trace of carbonaceous and other organic material formed as by-products of the operation. This steam passes to a stripping column in which the alcohol and ether are stripped out, leaving a residue of sulphuric acid and water of about 50 percent strength. The alcohol and ether vapors from the top of this stripping column pass to a scrubber in which they are washed with water or caustic. Steam is introduced into the base of this scrubbing in order to recover alcohol from the descending scrubbing liquor.

Vapors leaving the scrubber are condensed and cooled for purification and concentration in the finishing section of the unit. In this section the feed is first introduced into an ether column in which the ether is stripped from the dilute alcohol, concentrated in the top of the column, and back washed with water to remove final traces of alcohol. The ether-free alcohol from the base of the stripper passes to a rectifying column where the concentration of alcohol is raised to approximately 95 percent on a volume basis. Oils are drawn from an intermediate point in the column. The overhead product may then be further concentrated by stripping out additional heads materials at high proof.

Acid from the base of the stripping column passes to a reboiler in which the acid concentration is raised to approximately 70 percent, and vapor is provided for use in the base of the stripping column. This acid is then heated to accelerate carbon formation in the residual organic matter, oil material is decanted, and carbon is filtered out. The acid then passes to the acid concentrators in which water is removed by distillation under high vacuum, using Dowtherm (diphenyl and diphenyl oxide) as a heating medium. The concentrated sulphuric acid (94.5 percent) can then be re-used for the absorption.

Conclusions

The manufacturing costs of producing synthetic ethenol from ethanol under American conditions are less than $0.10 per gallon. This would include fixed charges, plus operating costs such as labor, maintenance, and utilities. Credits for the H_2SO_4 and ethyl ether have also been allowed in the above figure. The ethylene cost, which should be added, will depend upon the specific conditions involved. If by-product ethylene is used, the costs would vary to a large extent in accordance with the bookkeeping methods. In the case of a cracking unit directed toward organic chemical production, if crude oil were to cost only $2 per barrel, and if iso-propanol and other chemicals are made as joint products, the total cost of the ethanol would be less than $0.15 per gallon. It is believed that the total costs of making alcohol by direct catalytic hydration will be of the same order. These costs compare very favorably with the costs of producing ethanol from grain and of producing ethanol from grain and from blackstrap molasses. The prices of these agricultural commodities vary with the business cycle, but in general the price of grain far exceeds the necessary limit for economic production of industrial alcohol in countries where petroleum is available, while molasses could only serve as a marginal source during periods of low sugar prices. However, national and international considerations sometimes may modify these remarks, even in countries which have abundant petroleum resources.

The production of synthetic alcohol is another example of the application of sound chemical engineering techniques to the practical solution of an important industrial problem. The difficulties encountered in the numerous operations involved are indicative of the complexities of modern industrial processes which the chemical engineer has to design and operate. Similarly, the recovery of the by-product alcohol from synthol (Fischer-Tropsch) plants which at present gives rise to some problems, will be solved, making it an important process for organic chemicals manufacture.

* * *

A NEW TYPE . . .

Compressor Building For Gasoline Plant

By C. C. PRYOR
Associate Editor

Figure 1—Magnolia Petroleum Company's Vanderbilt, Texas, gasoline plant in the West Ranch field. The compressor building and suction and discharge headers are to the right of the line of process towers. Cooling water circulating pumps can be seen in the plant yard at the end of the compressor building.

DESIGN work on structures for supporting and housing compressors in gasoline, cycling, and repressuring plants has resulted in construction of a wide variety of types of foundations and buildings. Compressors, being quite heavy and subject to severe vibration problems, usually have been placed on foundations approximately at grade elevation. This arrangement worked well in installations where compressor suction and discharge pipe sizes were rather small and few stages were required. Piping was ordinarily placed in trenches below grade level, or supported overhead. The trend, however, in sizing suction and discharge piping has been toward larger diameters. This has come about because of attempts to obtain smoother flows to remove difficulties with piping vibration and gas measurement, and it has become more or less a frequent practice to use large surge volume tanks immediately adjacent to compressor inlet and outlet connections.

This tendency toward larger pipe sizes and surge volume tanks has resulted in many difficulties in piping layout. In some cases where grade level compressor foundations have been used, pipe trenches have been so crowded that original design was almost impossibly complex, and the resulting layout so

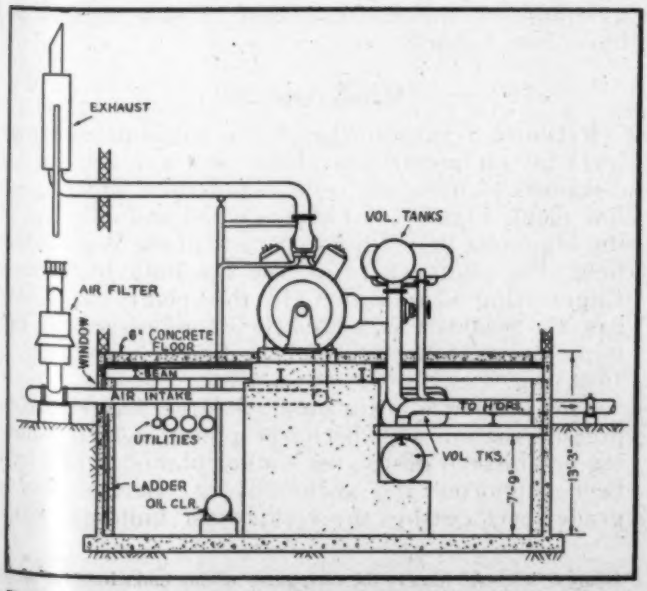


Figure 2—Cross-sectional sketch of the compressor foundation and basement, which shows relative location of equipment. Note that location of equipment provides easy accessibility.

THE physical layout of compressor installations possibly plays a greater part in providing good operating convenience and simple, low-cost maintenance than any other comparable mechanical equipment in the plant. In this account of one satisfactory design worked out to eliminate a number of the difficult problems encountered in conventional plants, it is anticipated that the reader will find several suggestions of value for future use.

Figure 3, above—Prefabricated manifolds of welded pipe for the engine air-intake are imbedded in the engine foundation block as shown in this picture of the block form under construction.

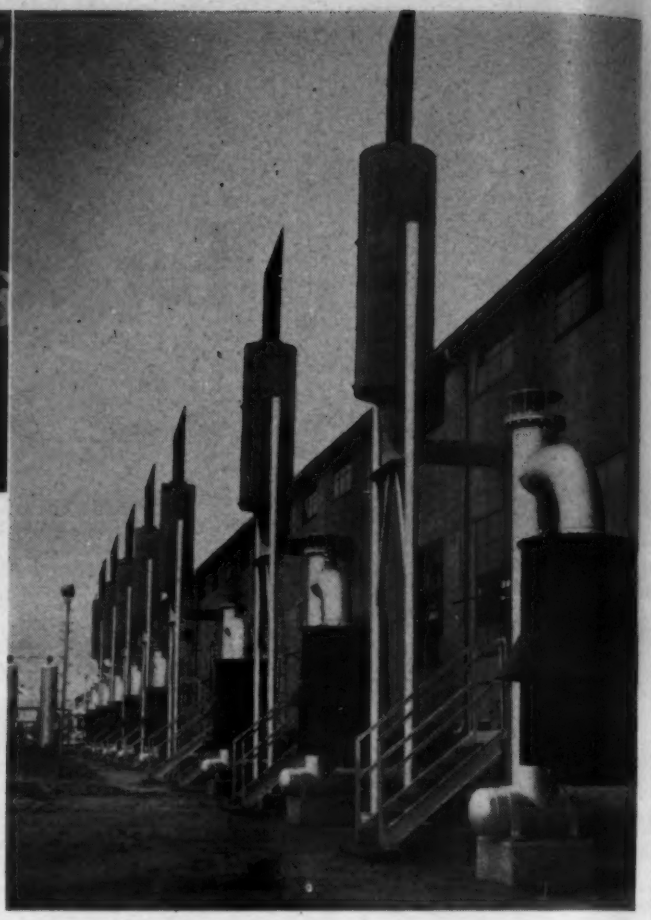


Figure 4, above—Air filter lines are brought through the basement wall under the building floor above grade level to the rear of the building. Lines run from the top of the engines through the wall to the exhaust silencers as shown in the photograph.

Figure 5, left—Rear of compressor engines showing insulated exhaust lines, and jacket water and fuel gas lines.

Figure 6, left—All lines to the compressor volume tanks from the suction and discharge headers in the plant yard are clamped to concrete piers and enter the compressor building above grade level and below the floor level.

crowded that many of the lines would be inaccessible for inspection or maintenance purposes. In addition to the layout problems noted above, one serious problem exists in plant operation where inaccessible pipe trenches are located below grade level to serve as traps for the accumulation of slop liquids and hazardous vapors.

Vanderbilt Plant

Recently a variation from the conventional grade level layout mentioned above was evolved for the compressors installed in the Vanderbilt, Texas, gasoline plant, Figure 1, which is owned and operated by the Magnolia Petroleum Company in the West Ranch field. The plant was designed and built by Hudson Engineering Corporation. In this plant, casinghead gas, 27,500,000 cubic feet daily, is compressed through four stages form 10 inches mercury vacuum to 1000 psig.

This particular plant employs large diameter compressor suction and discharge piping, which following established design for similar plants, would have been supported and anchored on concrete piers at grade level outside the compressor building. Large

Figure 7—Installation of volume tanks for compressor discharge is above cylinders. Note that low pressure tanks are across two low pressure cylinders of engines. Each compressor is provided with five cylinders compressing gas in four stages from 10 inches mercury vacuum to 1000 psig.

volume tanks were to be installed on the inlet and outlet lines from the compressors.

The design worked out to meet this layout problem was based on the elevation of the compressors to a level several feet above grade. The compressors are supported on individual pedestals which are extensions of their normal foundation blocks. The operating floor is of course at its usual position with relation to the compressors and is carried on a floor beam system supported by the compressor pedestals and the outside walls of the compressor building. Concrete work required for a trench system often was about as expensive as a basement type design because of the complex shape and framework design time, et cetera.

The basic foundation for all compressors and the compressor building is a reinforced concrete mat 280 feet in length, 35 feet wide and 18 inches thick. Top surface of the mat constitutes the basement floor and is some 10 feet below grade level.

The sketch of Figure 2 is a cross-sectional view showing the mat and concrete blocks for the engines and the large basement that resulted from this type of construction.

The individual concrete blocks, almost 14 feet high, 24 feet long, and 11 feet wide were poured for each of the eight compressor engines installed in line the length of the building. There is a space of nearly six feet between blocks in the basement. The base of the blocks is approximately 14 feet wide. Reinforced concrete walls nine inches thick were built around the mat to the same height as the compressor engine blocks. I-beams from the walls to the blocks support a reinforced concrete floor slab designed for a live load of 400 pounds per square foot.

The compressor engines were installed on the foundation blocks, the floor slabs poured, and the corrugated sheet steel compressor building erected. As shown in Figure 2 and in the photograph, Figure

Figure 8
Rear view of foundation blocks in the basement. Pipe lines for fuel gas, engine jacket water, and lubricating oil coolers are suspended from I-beams supporting the building floor. Lubricating oil coolers are at lower left. Not shown, but in the basement also, are Lubricating oil filters. There is a space of nearly 6 feet between the ends of each concrete block.

Figure 9
Expansion loops connect the lines from the compressor volume tanks to the suction and discharge headers in the plant yard, which are mounted on concrete piers.

3, the welded air intake manifold for each engine was prefabricated from pipe and is in the concrete foundation block beneath the engine. This method eliminated running the manifold above the engines or alongside the foundation blocks, which would have increased the amount of suspended piping in the basement space. The connecting line to the intake filter outside the building crosses from the foundation block to the back wall of the building through an opening in the concrete beneath the floor slab and other piping that runs the length of the building. These lines are also above ground and are clamped to concrete supporting piers as shown in Figure 4. Exhaust silencers for the engines are behind the building in line with the air intake filters. Lines from the engines to the silencers are above the engines at an elevation high enough to be out of the way of operating personnel, Figure 5.

Piping

The floor level of the building is approximately five feet above ground level, so that piping from the suction and discharge headers in the plant yard for the four compression stages enters the basement through the concrete wall (Figure 6) and is laid above the ground and clamped to concrete piers for accessibility and maintenance, and as a further aid in reducing line vibration. Since the compressors are in a single line, all piping to the headers is on one side of the building. Volume tanks, used to reduce pulsation, for the inlet lines are in the basement directly beneath the cylinders to which they are connected. This permits maximum utilization of the tanks by providing header pressures at the compressor cylinders. Each compressor has five cylinders, two of which are used for the first stage. The discharge volume tanks for these cylinders, as shown in Figure 7, are above the compressors across the cylinders and are supported by the connecting piping.

Volume tanks in the basement are installed in a manner similar to those above the compressors. The small tank on the high pressure cylinder is beneath the slab floor as near as practical to the cylinder; whereas the larger tanks on the cylinders operating at lower pressure are nearer to the step at the base of the foundation block. This arrangement of the tanks is shown in the sketch of Figure 2.

Utilities

Figure 8 is a photograph of the rear of the engine foundation blocks in the basement and shows the engine jacket and lubricating oil cooling water lines and fuel gas lines suspended from the slab floor supporting beams. They are accessible for inspection and maintenance. Although not shown in the photographs, oil filters are also installed in the basement. Installing these items of accessory equipment in the basement permits all-weather maintenance. Centrifugal pumps, driven by explosion-proof motors, supply cooling water circulation, and are located at one end of the building in the plant yard, Figure 1.

Design of the plant is such that the compressor house and basement can be extended readily for additional compressors, and blind flanges can be removed from the end of present headers for easy extension. As the headers are all anchored, expansion loops (Figure 9) were used to connect them to the compressors; thus achieving piping clearance and flexibility. The field suction line, 42 inches in diameter, is the largest known to be in use in a gasoline plant.

Work Continuing On Shell's Plant In West Venezuela

Welding Pipes at Refinery Which Shell Company of Venezuela is building at Cardon on the Peninsula of Paraguana.

SHELL Company of Venezuela has a modern refinery under construction at Cardon on the peninsula of Paraguana on the eastern shore of the Gulf of Venezuela. Though there is an entire lack of water on this barren desert, the site is on deep water, affording safe berths for deep draft ocean tankers and a clear run to U. S. and European ports.

Started in 1945

Since 1945, when the project was started, Shell engineers have installed four modern units for distilling sea water for drinking purposes, and additional water for industrial and sanitary purposes is brought in by tanker and pumped from tanks to refinery and living quarters. Now, under joint consideration by Shell and Creole Petroleum Corporation, which is also constructing a refinery nearby, is the construction of a water pipe line from the town of Coro on the Venezuelan mainland, 69 miles away.

Current reports from the Shell staff at Cardon indicate that rapid progress has been made in transforming the desolate desert area into the nucleus of a well-planned model community. Already completed, after 2½ years, are one of two contemplated churches; 325 buildings which already accommodate 6500 workers; temporary electric light and power plants; sanitation and water systems; a network of over 150 miles of roads; two of three projected school buildings; a barber shop, beauty parlor, a cinema and theater for 1200 persons, and a cemetery and its chapel.

Many Conveniences

In process of construction are a large modern hospital, mess halls, clubs, laundry, a restaurant, a commissary, a number of retail stores, a bank, a post office and many recreational facilities, such as a tennis court, baseball park, cricket pitch, a nine-hole golf course, and the improvement of several natural bathing beaches. Plans have been made for the creation of a model community to house 10-12,000 inhabitants.

Progress Despite Difficulties

Apart from the community aspect, considerable progress also has been made in the construction of the refinery and its many adjunct facilities, in spite of the difficulty of obtaining essential materials and equipment. Already completed are 12 large storage and 15 smaller tanks, several store material buildings, extensive yard facilities, machine, electric, instrument, welding, carpenter and repair shops, a central garage building; also automotive equipment maintenance shops, a rock quarry with rock-crushing plant, a concrete block-making plant, a pipe shop, a welding school, and two barge slipways on which fourteen 200-ton barges have been built.

Under construction are many additional tanks, a crude distilling plant, a vacuum distillation plant, a redistilling unit, a cracked products separation plant, two thermal cracking plants, and a thermal reforming plant.

The first of two contemplated reinforced concrete jetties is already completed, and is now being used chiefly for the unloading of supplies from abroad. These jetties are each capable of handling simultaneously two shallow-draft tankers and two ocean-going tankers of the largest capacity. The second jetty was due for completion early in 1948.

Construction of Chimney at Shell's New Refining Plant on Eastern Shore of Gulf of Venezuela. Jetty No. 1, 1440 Feet Long and Already Completed, in Background.

Between the Jetties

In the area between the two jetties, several plants and installations are being erected along the ocean front: a large electric power plant, which will supply power and light for the refinery proper and for the extensive living area; a salt water pumphouse which will provide cooling water for the refinery plant; a boiler plant containing both high pressure and low pressure boilers; a fifth large salt water distilling plant, and a boiler-feed water treating plant.

No completion date has been set for the Cardon refinery, which will have a daily production capacity in excess of 40,000 barrels of all types of refined products.

Applications of Thermodynamics to Hydrocarbon Processing

Part X—PSEUDO-CRITICAL FOR MIXTURES

WAYNE C. EDMISTER, Hydrocarbon Research, Inc., New York

THE computation of vapor densities in the temperature and pressure ranges encountered in most hydrocarbon processes necessitates recognizing deviations from the ideal gas law. This may be done by the compressibility factor, the volume residual, or by an analytical equation of state, employing generalized methods with reduced pressure and temperature parameters. Thermodynamic equations may use the same generalized methods.

For pure components the reduced conditions are calculated from the critical point and the resulting generalized correlations and derivations are of sufficient accuracy for practically all process engineering calculations. Thermodynamic properties of mixtures may be computed from the properties of the individual components by combining the partial quantities for the components in the proper fashion. Another method is to use an equation of state with constants from combining component constants. Both of these methods are lengthy for multi-component hydrocarbon systems so it is desirable to use generalized methods where possible. In applying generalized methods to mixtures, average critical conditions instead of true critical conditions must be used in computing the reduced temperatures and pressures for the system. Such an average critical temperature and pressure was proposed in 1936 by Kay[4] and is known as the pseudo-critical point. By using reduced conditions computed from pseudo-critical as the dependent variables, the P-V-T data fit the same generalized correlations that had been previously developed from the P-V-T data of pure components. This is a very useful and important concept in the applications of thermodynamics to hydrocarbons.

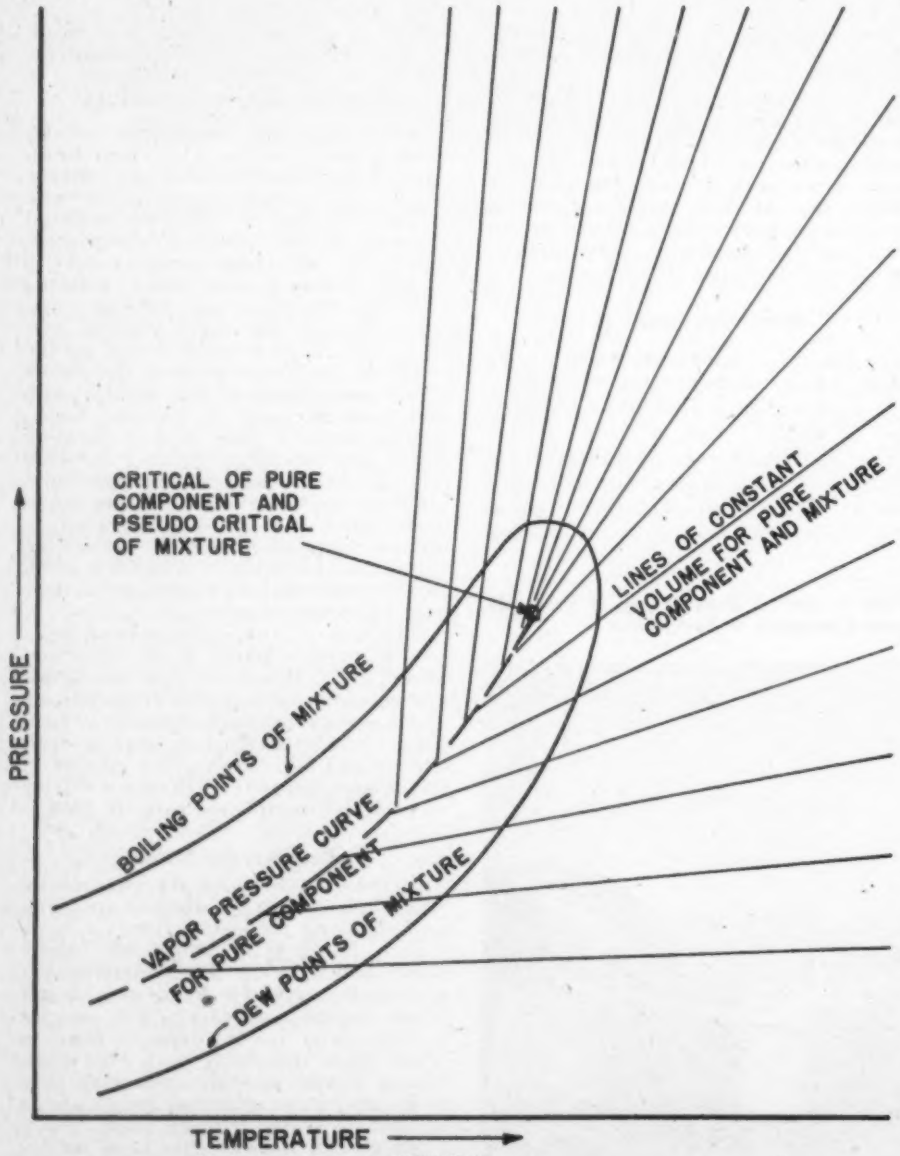

CRITICAL OF PURE COMPONENT AND PSEUDO CRITICAL OF MIXTURE
LINES OF CONSTANT VOLUME FOR PURE COMPONENT AND MIXTURE
PRESSURE
BOILING POINTS OF MIXTURE
VAPOR PRESSURE CURVE FOR PURE COMPONENT
DEW POINTS OF MIXTURE
TEMPERATURE

FIGURE 21
Pressure-Temperature Plot for Typical Single and Multicomponent Systems, Illustrating Pseudo-Critical.

The pseudo-critical is ordinarily used with graphical correlations such as the compressibility factor or the volume residual. It may also be used with analytical equations of state thus applying equations of state developed for pure components to mixtures by means of reduced conditions calculated from the pseudo-critical temperature and pressure for the entire mixture.

As will be pointed out in this chapter, the combination of empirical equation of state constants to obtain constants for a mixture is analogous to the combination of the critical constants of the components to obtain a pseudo-critical

constants for the mixture. Thus the pseudo-critical concept is no more empirical than the equation of state method for mixtures.

In addition to being used to compute the vapor densities for hydrocarbon mixtures, the pseudo-critical may be used to estimate the isothermal pressure corrections to the entropy and the enthalpy from generalized functions that are based upon thermodynamic derivations made for pure components.

Two other applications of lesser importance for the pseudo-critical are the use of it as a means of correlating the true critical for mixtures, and also the use of it to estimate the fugacity of components in gaseous mixtures where neither ideal gas nor ideal solution laws apply.

In studying the pseudo-critical concept, it is of interest to have in mind the five different reference points that will be utilized in applying thermodynamics to hydrocarbons. They are as follows: a) true critical, b) pseudo-critical, c) focal point, d) cricondentherm, and e) convergence pressure. Of these five reference points, the true critical is used in all generalized computations for pure components whereas the pseudo-critical is used for generalized computations for mixtures. The focal, cricondentherm, and convergence pressure points are used in multicomponent hydrocarbon phase equilibria computations.

IN THE application of thermodynamics to hydrocarbon processing, one of the most useful tools is the pseudo-critical concept. The pseudo-critical conditions are used in calculating reduced pressures and temperatures for hydrocarbon mixtures in the same way the true critical conditions are used for pure components, thus fitting both single and multicomponent systems into the same generalized correlations. The pseudo-critical temperature and pressure are hypothetical properties of hydrocarbon mixtures, the values depending upon the composition of the system. Pseudo-critical temperatures and pressures are calculated from the true critical temperatures and pressures of the components by empirical equations or, alternately, they are found directly from gravity and boiling point characteristics of the mixture by means of graphical correlations.

Although the pseudo-critical concept was proposed more than a decade ago, its potentialities have not yet been fully realized. In addition to the computation of vapor densities for hydrocarbon mixtures, pseudo-criticals may be used in estimating the isothermal pressure corrections to the entropy and enthalpy for multicomponent systems. Also, pseudo-critical may be used to predict the true critical of a mixture and to estimate the fugacity of a multicomponent vapor.

The Critical Point

In Part VII (PETROLEUM REFINER, January, 1948), the P-V-T relationships for single and multicomponent systems were illustrated by diagrams showing three different methods of plotting. On each of the resulting six diagrams, the critical point is indicated by a small circle at the top of the two phase boundary curve. The critical point of either pure component or mixture is the juncture of the saturated liquid and saturated vapor lines. On P-T scales these two curves coincide to form a single vapor pressure line for a pure component. On P-V and V-T scales, the saturated liquid and vapor lines for pure components bound two phase areas. In all three plots, namely P-T, P-V and V-T, for mixtures the saturated liquid and vapor lines bound two phase areas, intersecting at the critical point.

Thus, it can be seen that the critical point is always on the two phase border curve for both single and multicomponent systems. Lines of constant quality, i.e., percent vaporized, drawn in the two phase area converge at the critical point for both pure components and mixtures.

The critical point may be determined experimentally for either mixed or pure systems. Observed critical data are available in the technical literature. In Table 13 of Part VII, critical constants are given for 32 hydrocarbons.

The estimation of critical conditions for hydrocarbon mixtures from the values for the individual components is relatively easy and straightforward as far as critical temperature is concerned, but very difficult and complex as far as the critical pressure is concerned. The critical temperature of a mixture is the weight average of the critical temperatures of the components. The critical pressure, on the other hand, is no simple average of the component critical pressures, and in fact may be greater in value than the critical pressure of any of the components of the systems. Mixture critical pressures are dependent upon the width of the fraction and upon the distribution of components.

Several attempts have been made to correlate the critical conditions of hydrocarbons and petroleum fractions.[2, 5, 8, 11, 12] Most of this work has been for petroleum fractions as defined by the A.S.T.M. distillation and gravity, which is necessary because of the accepted practices for characterizing petroleum fractions. It is, nevertheless, awkward to have critical conditions as a function of °API and ASTM when it comes to making thermodynamic computations for the fraction, particularly for those mixtures that are just too heavy to define conveniently by the component analysis method.

Watson and Nelson[12] correlated critical temperature as a function of molal average boiling point, and found critical pressure from a Cox chart by a graphical method employing a focal point which was located on an average boiling point line beyond the critical temperature a distance of one-fifth of the distance between the atmospheric boiling point and the critical point, also measured along the average boiling point line.

Smith and Watson[8] correlated true and pseudo-critical temperatures on the same plot, using average boiling point (weight average for true critical and molal average for pseudo-critical) and gravity. These authors point out that the lack of both true critical and pseudo-critical data for the same mixtures makes the development of correlations difficult. The situation has not changed much in the ten years since Smith and Watson made this observation. However, they developed two correlations for the pseudo-critical pressure, one as a function of mean average boiling point and gravity, and the other as a function of molecular weight and gravity. Also presented by these authors is a relationship between the ratios of true to pseudo-critical pressure and true to pseudo-critical temperature. By means of this relationship, observed true critical pressure data were used in developing the pseudo-critical pressure correlations.

In 1942, Kurata and Katz[5] presented experimental critical data on eleven volatile hydrocarbon mixtures. These authors also present correlations for true and pseudo-critical. These correlations use in addition to the calculated (from composition and individual criticals) pseudo-critical temperature and pressure, an "equivalent" molecular weight in determining the

critical temperature and pressure of volatile hydrocarbon mixtures.

From the above discussion, it can be seen that the relationship between true and pseudo-critical is complex.

Pseudo-Critical Phenomenon

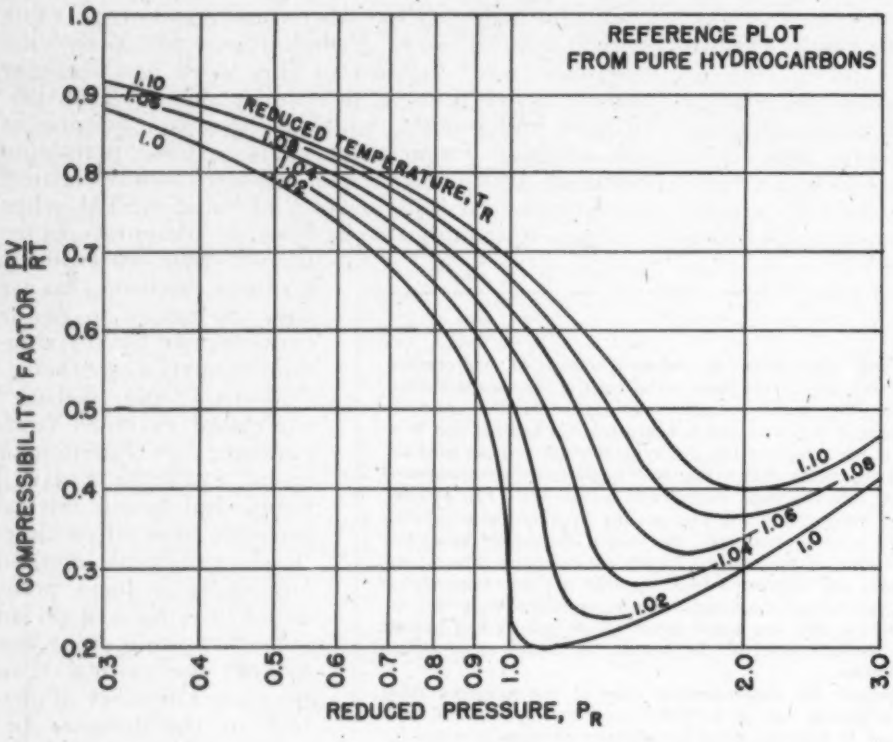
In the study of thermodynamics, the pseudo-critical point is of interest because it simplifies the applications of the theory to hydrocarbon mixtures. In this connection, it is of interest to note that modified criticals have been applied to pure components to improve the accuracy of generalized thermodynamic correlations. This involves using empirically modified critical constants for those components where such changes would improve the reduced correlations. This is what Newton[7] did in 1935 for helium, hydrogen and neon to make the compressibilities of these gases fit a compressibility vs. reduced pressure and temperature plot prepared for other gases, including hydrocarbon, organic, inorganic, etc., gases.

In order to avoid confusion, the term *pseudo-critical* will be used only for mixtures. Where critical points of gases are adjusted to improve the P-V-T correlation, the revised pure component critical values will be called *modified* critical conditions.

By 1936 several authors had correlated the compressibility factors, or deviations from ideal gas laws, for pure hydrocarbons with reduced temperature and pressure. Kay[4] observed that the data for complex mixtures deviated widely from the pure hydrocarbon plots when plotted by the same generalized method using the critical conditions of the mixture. This deviation was explained by Kay as being due to the improper selection of the critical reference point, which prior to this time had been the true critical for the mixture. To remedy this situation, Kay proposed the pseudo-critical.

The pseudo-critical concept was illustrated by Kay in his 1936 article by a drawing similar to Figure 21. On the pressure-temperature scales of this plot are shown: a) the vapor pressure curve of a pure component, b) the phase (liquid-vapor) border curve of an equivalent mixture, c) lines of constant volume for both single and multicomponent systems, and d) the critical of the pure component and pseudo-critical of the equivalent mixture.

In proposing the pseudo-critical and describing this diagram, Kay explained that the lines of constant volume for the mixture and pure component would coincide if both single and multicomponent systems were plotted on reduced temperature and pressure scales, using the common reference point indicated. It will be noted that the isometrics for both systems converge at this point. The true critical of the mixture is located on the dome of the border curve. The fact that reduced isometrics do not coincide for different pure components on the P_r vs. T_r scales



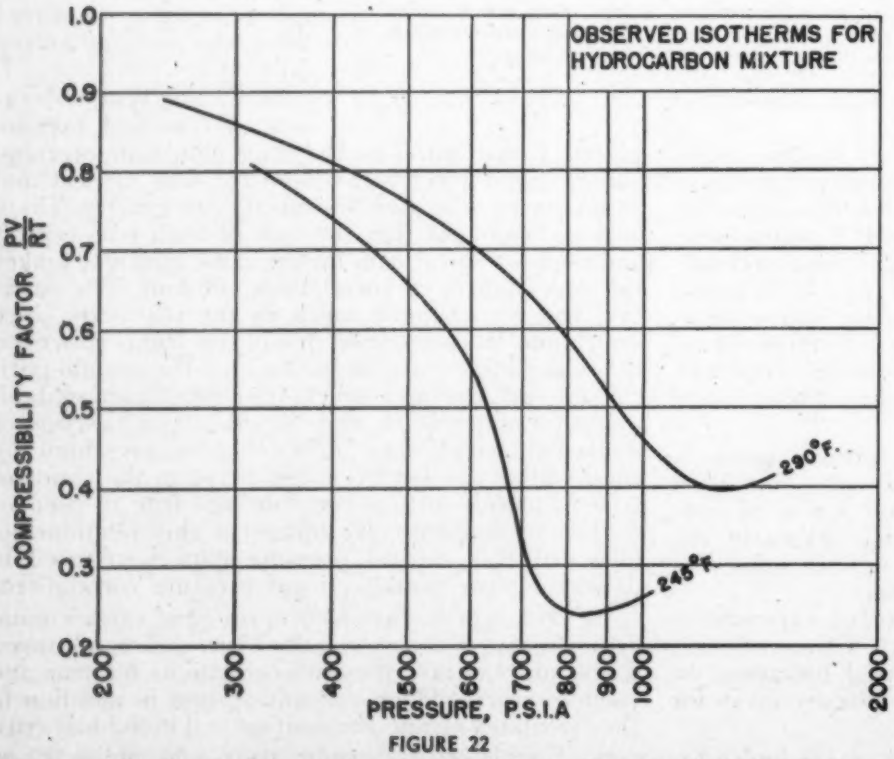


FIGURE 22

Compressibility Plots for Reference Component and Mixture, Illustrating Method of Finding Pseudo-Critical.

does not invalidate the pseudo-critical method.

From the phase diagram in Figure 21, it is clear that the pseudo-critical of a mixture differs from the true critical of the mixture by an amount depending upon the boiling point and range of the mixture, etc. The critical conditions are definite properties of a mixture whereas the pseudo-critical conditions are hypothetical properties. The true critical conditions may be measured experimentally in the laboratory whereas the pseudo-critical must be found by indirect means.

On the P-T phase diagram for a mixture, the lines of constant percent vaporized in the vapor-liquid region converge at the two phase critical point, which is also the juncture of the saturated liquid and saturated vapor or *bubble* and *dew* point curves.

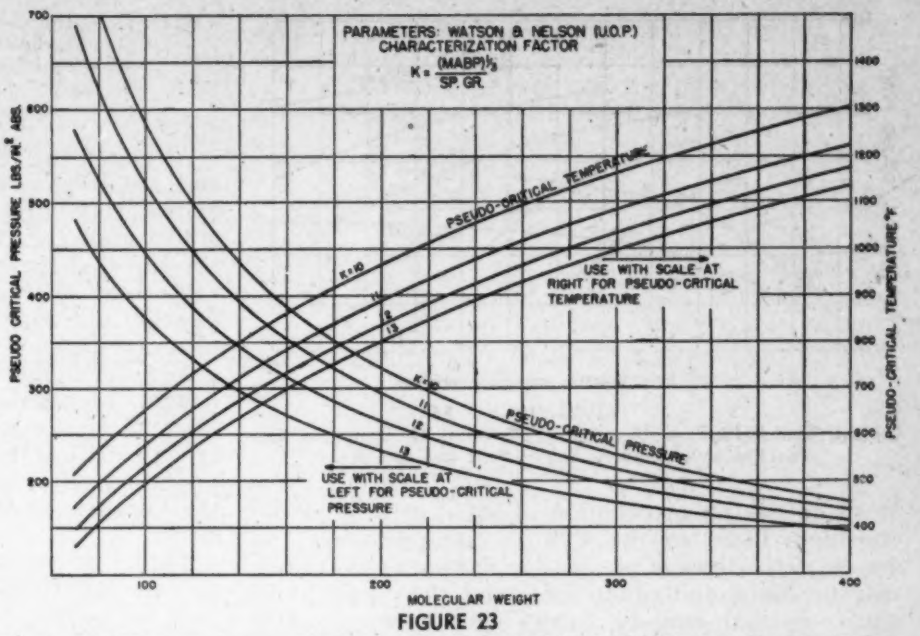


FIGURE 23

Pseudo-Critical Temperature and Pressure as Function of Molecular Weight and Characterization Factor.

[Redrawn from Figure 8, page 1019, of W. B. Kay's "Density of Hydrocarbon Gases and Vapors," Ind. Eng. Chem., 28, p. 1014 (1936).]

Determination of Pseudo-Critical

As shown in Figure 21, the use of an average or pseudo-critical point for mixtures will permit fitting P-V-T of mixtures to pure component correlations. In order to make use of this concept, there must be a method of predicting the pressure and temperature conditions of the pseudo-critical point.

The pseudo-critical temperature and pressure of a mixture cannot be measured directly because it is a hypothetical property depending upon P-V-T behavior. Accordingly, the pseudo-critical conditions must be determined indirectly. Kay found the pseudo-critical conditions for eleven hydrocarbon mixtures by making experimental P-V-T determinations, and from these data finding, by means of plots, the pseudo-critical temperatures and pressures required for each mixture to make the observed P-V-T data fit on a compressibility plot such as Figure 19 of Part VIII (PETROLEUM REFINER, February, 1948). These pseudo-critical conditions were then correlated and a method presented for estimating the pseudo-critical for any hydrocarbon mixture from its composition or its ASTM distillation characteristics. The correlation and prediction of the pseudo-critical temperatures and pressures will be discussed later.

Figure 22 illustrates the way Kay found these pseudo-criticals. First, a reference PV/RT plot was prepared from pure hydrocarbon data on transparent semi-log paper, using reduced temperatures and reduced pressures. The upper plot in Figure 22, which is a replot of Figure 19, is this reference plot. Second, the observed PV/RT data for the mixtures were plotted on the same kind of graph paper, but using actual temperatures and pressures instead of reduced conditions. The lower plot of Figure 22 is one of these observed data plots. Eleven plots like this one were made, one for each mixture studied. Only those temperatures for which experimental data were taken were included on these plots.

The observed isotherm plots were then matched up with the reference plot, one at a time, so that the PV/RT scales coincided and the observed mixture isotherms coincided with the reduced isotherms on the reference plot. This was done against a window pane by first lining up the PV/RT scale and then shifting the two plots along the pressure scale until the isotherms coincided.

When the two graphs were thus matched up, the values of the pseudo-critical temperature and pressure required for the mixture to give reduced temperatures and pressures corresponding to the reference plot were determined by the matching position. Logarithmic pressure scales of the same cycle length were used for both plots because this determined pressure scales that only differed by a constant amount, which made it possible to match up the isotherms. This is easily understood by considering the relation:

$$\log P_r = \log P - \log P_c$$

Where: P_c is the critical or pseudo-critical pressure.

When the isotherms on the "reference" and "unknown" plots are matched, the pseudo-critical pressure and temperature of the mixture are found from the readings of both plots at coinciding points.

From the values of the pseudo-critical temperatures and pressures thus found by these graphical analyses of observed mixture compressibilities, Kay developed two correlations for estimating the pseudo-critical conditions for multicomponent hydrocarbon systems. One of the correlations proposed by Kay uses the mixture composition and the criticals of the individual components. The other correlation uses the Watson-Nelson (UOP) characterization factor and the molecular weight.

The above describes the graphical method used by Kay in finding the pseudo-critical conditions for eleven hydrocarbon mixtures. A mathematical method could be used for finding the pseudo-critical. For pure components the critical isotherm on P-V scales has an inflection at the critical point so that

$$\left(\frac{\partial P}{\partial V}\right)_T = 0$$

$$\left(\frac{\partial^2 P}{\partial V^2}\right)_T = 0$$

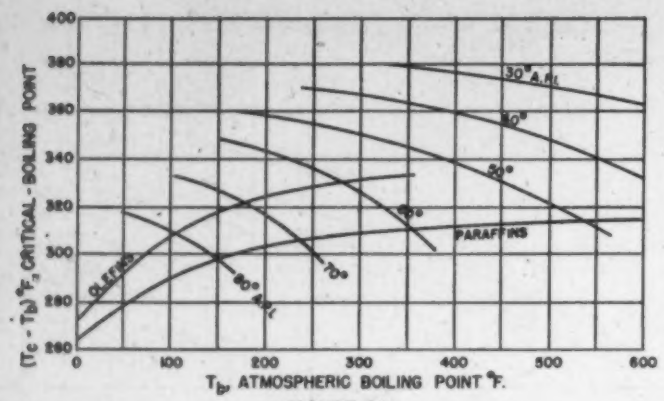


FIGURE 24
Critical Temperatures of Hypothetical Hydrocarbon Components as Function of Atmospheric Boiling Point and Gravity.

With an analytical equation of state with empirical constants fitted to the P-V-T data, the first and second derivatives of pressure with respect to volume may be found and equated to zero, thus giving two equations that may be solved simultaneously with the original equation of state to find the values of pressure, volume and temperature at the critical point. This is the way the equations giving the van der Waals constants a and b as functions of T_c and P_c were obtained (see equations 165 and 166 in Part VIII).

For a mixture, the pseudo-critical isotherm would have an inflection point if drawn through two phase field, which it could be from an equation fitted to experimentally determined P-V-T data. Figure 15 in Part VII shows the difference between the P-V plots for single and multicomponent systems but does not show the critical and pseudo-critical isotherms nor the points of inflection thereon. However, it follows from the nature of the equations that such a point exists.

In stating that an inflection point occurs on the pseudo-critical isotherm for a mixture, it is assumed that the use of pseudo-critical temperature and pressure as the reference point for computing reduced conditions makes it possible to correlate both single and multicomponet compressibilities in one generalized correlation.

Theoretically, the above could be done but it would not be very practical because of time required for fitting the equation of state to the experimental P-V-T data, differentiating this equation to get the first and second derivatives and then solving these three equations simultaneously. Also, most equations of state are not rigorously accurate in all ranges, particularly in the critical region. In applying equations of state, it is customary to make residual corrections or to employ graphical methods in the critical region. It appears that the method used by Kay is the most practical, accurate and consistent procedure for determining the pseudo-critical conditions of a mixture.

Estimation of Pseudo-Critical Conditions

For the light hydrocarbon mixtures, which are usually defined by component composition, Kay proposed obtaining the pseudo-critical conditions as the summation of the products of the mol fraction and the critical points of the individual components. Expressed mathematically

$$T_{pc} = \Sigma x_i T_{c_i} \quad (193)$$

$$P_{pc} = \Sigma x_i P_{c_i} \quad (194)$$

where: T_{pc} = the pseudo-critical temperature of the mixture
P_{pc} = the pseudo-critical pressure of the mixture
x_i = mol fraction of the i^{th} component
T_{c_i} and P_{c_i} = true critical conditions of the i^{th} component.

Although the relationships given by equations 193 and 194 were obtained from observations on light hydrocarbon mixture P-V-T data and are most generally used for the light hydrocarbon mixtures, it appears that these relations should be equally applicable to all hydrocarbon mixtures for which the composition and critical conditions of the individual components are known.

For the higher boiling mixtures, the exact component compositions are usually not known. In some cases there are components for which there are not reliable critical data. These two situations make empirical correlations employing average properties of the mixtures desirable for the pseudo-critical conditions.

Figure 23 is the correlation proposed by Kay. In this chart both pseudo-critical pressure and temperature are plotted against molecular weight for lines of constant characterization factor. This correlation was developed by Kay from the pseudo-critical data he obtained plus the critical data on pure hydrocarbons. The pure component data were included because this is the limiting case for the correlation and because using these additional data gave more points and made it possible to develop a more complete correlation. In constructing this plot, Kay used modified scales that gave straight lines in order to permit extrapolating to the high molecular weight range. The smoothed and extrapolated data were then replotted to get Figure 23.

As originally proposed and applied, these two methods (equations 193 and 194 and Figure 23) were both used. For light gaseous hydrocarbon mixtures of known composition usually consisting of one to five carbon atom hydrocarbons, equations 193 and 194 were used to estimate the pseudo-critical temperature and pressure. For petroleum fractions of known gravity and ASTM distillation, Figure 23 was used.

When the multicomponent system in question contains both light gaseous hydrocarbons and a petroleum fraction (such as naphtha, kerosine, gas oil, etc.), the pseudo-critical conditions may be found in two alternate ways. If the amount of gas is small in proportion to the quantity of the petroleum fraction, Figure 23 may be used with the molecular weight and K factor for the entire mixture. In obtaining the latter, an ASTM distillation curve for the entire system may be constructed by correcting the ASTM of the petroleum fraction for the carrying effect of the light gaseous hydrocarbons.

Another procedure for estimating the pseudo-critical conditions of such wide mixtures is to find the pseudo-critical conditions for the light gaseous hydrocarbon mixture and the petroleum fraction separately, using equations 193 and 194 for the gaseous portion and Figure 23 for the liquid fraction, and combining these pseudo-critical constants by the relative molal quantities of the gaseous and liquid portions.

Pseudo-Critical in Generalized Equation of State

With the pseudo-critical concept the P-V-T data for mixtures can be correlated by graphical methods, such as the PV/RT plot used by Kay in developing the pseudo-critical. This suggests the possibility of using the pseudo-critical and generalized reduced

equations of state for mixtures. Much of the work that has been done on the development of generalized equations of state for pure components, where true criticals were used, is of interest in considering a similar type of equation for mixtures using pseudo-critical constants.

Maron and Turnbull[8] calculated Beattie-Bridgeman equation of state constants for various gases from the constants of a reference gas (nitrogen) and the critical temperatures and pressures of the gas in question and the reference gas. In converting any one of the constants from the reference gas value to the value for the other gas, the ratios of critical temperatures and pressures of the two gases are used as multiplying corrections in a manner determined by the character of the particular portion of the equation of state where the given constant appears. This last consideration determines the power to which these ratio of critical temperatures and pressures must be raised.

Maron and Turnbull also developed generalized equations for the thermodynamic properties by: a) obtaining a generalized equation for the activity coefficient by substituting the approximate explicit-in-volume form of the Beattie-Bridgeman equation of state into the activity coefficient equation, integrating and converting to reduced units, and then evaluating the constants, b) from this generalized activity coefficient equation the other thermodynamic properties are formulated.

Su and Chang[9] generalized the Beattie-Bridgeman equation of state in another way. These authors started out with the original Beattie-Bridgeman equation of state (equation 167 of Part VIII) written with reduced pressures, temperatures and volumes in place of actual P, T and V terms, and with five generalized constants in place of the original constants. The form of the equation was not changed but the terms were made dimensionless. In this rewritten generalized equation, the reduced volume was the actual volume divided by the ideal critical volume $V_{c1} = RT_c/P_c$ which gets around the uncertainty that exists about the values of the actual critical volume. The reduced constants were evaluated from known specific constants of various gases and compared. This method of evaluation involved using only the critical pressure and temperature and the empirical constants for a given gas.

Another method used by these authors in evaluating the reduced constants was to work direct from compressibility data, which constants were shown to agree very well with the average of those obtained by the first method for eleven lighter gases from argon through n-butane. The "generalized" constants obtained from different sets of "specific" constants do not vary widely for these lighter gases but higher

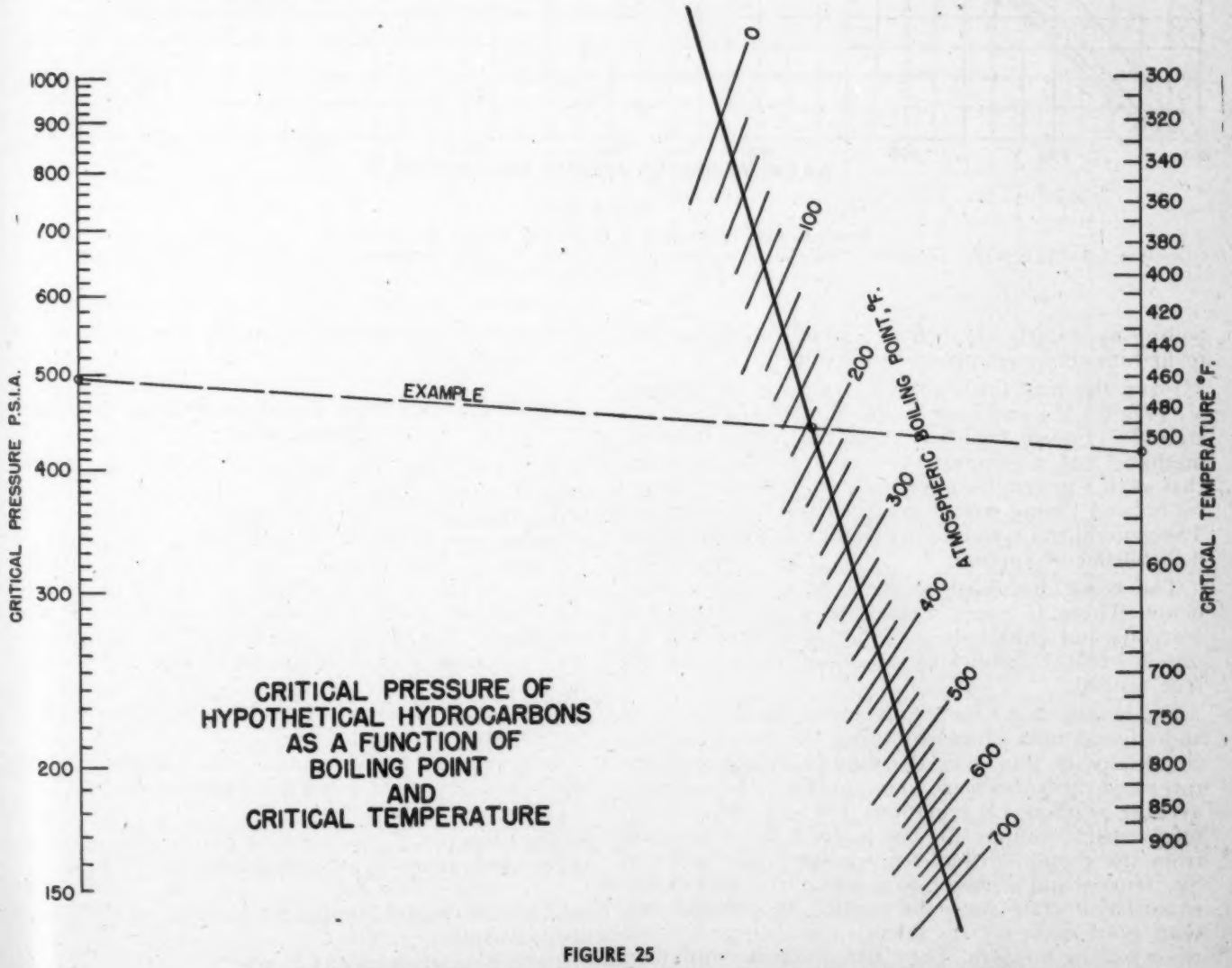


FIGURE 25
Critical Pressures of Hypothetical Hydrocarbon Components as Function of Atmospheric Boiling Point and Critical Temperature.

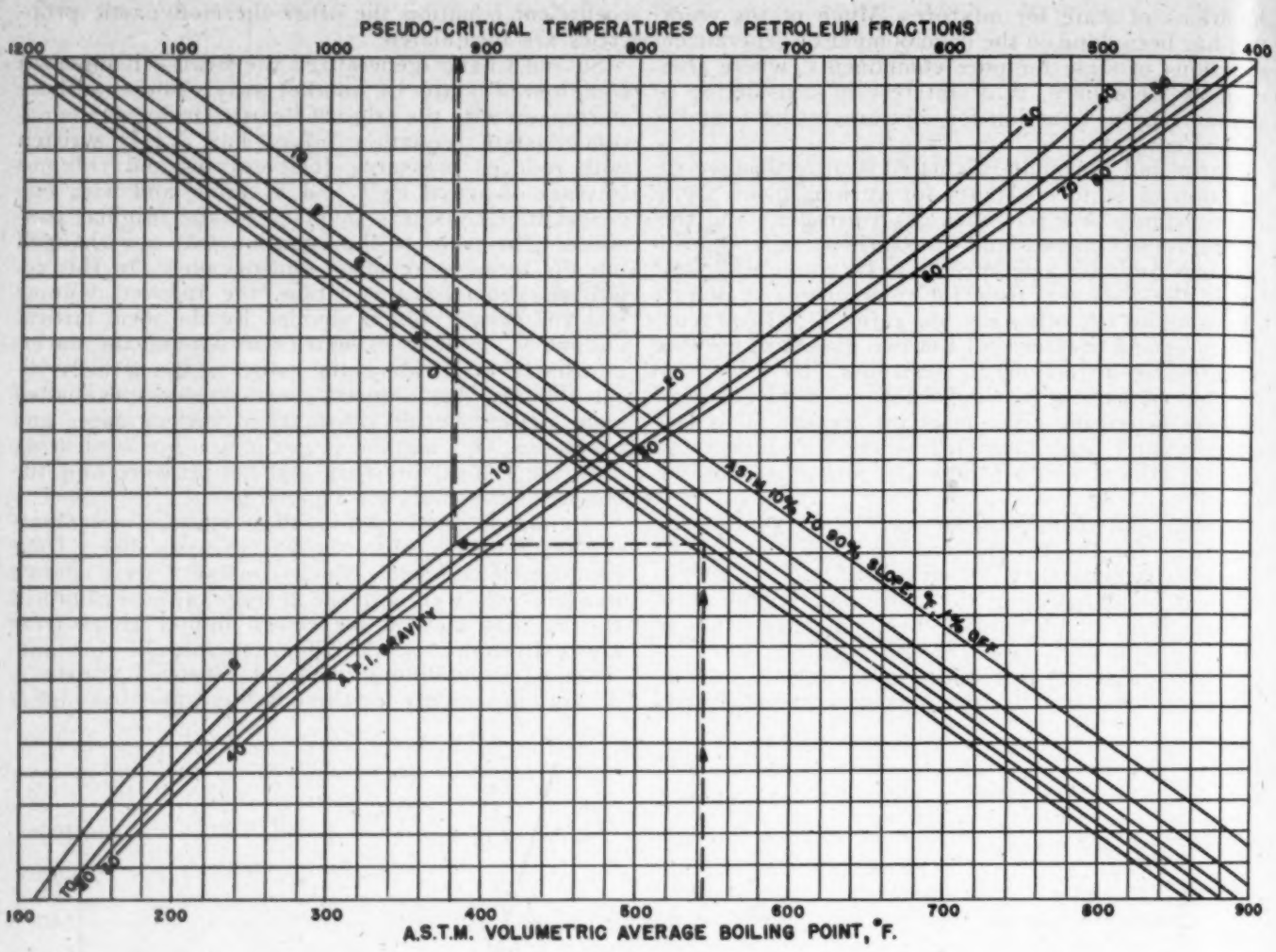


FIGURE 26

Pseudo-Critical Temperature as Function of V.A.B.P. and Gravity.

(Developed from W. B. Kay's correlation, Figure 23, by combining with molecular weight correlation, Figure 14 (Part VII), and definition of U.O.P. Characterization factor.)

boiling hydrocarbons, such as heptane, do not appear to fit into this generalization very well.

From the way the P_r vs. T_r for lines of constant V_r (Figure 17) and Log V_r vs. T_r for lines of constant P_r (Figure 18) plots show the spread between methane and n-pentane, it would not be expected that such a generalized equation of state as proposed by Su and Chang would apply to the hydrocarbons. The generalized activity coefficient method has better possibilities of success.

The work discussed above was on pure components. There is every reason to expect the same methods and equations to apply to mixtures if the pseudo-critical conditions are used instead of the true critical.

Su, Huang and Chang[10] extended the work of Su and Chang[9] to mixtures by using the pseudo-critical conditions. In this work the pseudo-critical temperatures and pressures were found by Kay's linear molal average method, i.e. equations 193 and 194, and the ideal critical volume for the mixture was computed from the pseudo-critical temperatures and pressure. Su, Huang and Chang[10] found that the generalized equation of state could be applied to gas mixtures with good accuracy by using pseudo-critical conditions for the mixture. They also show the similarity between this method of applying a generalized equation of state by the pseudo-critical with the usual method of combining the equation of state constants for a mixture.

Pseudo-Criticals from Equation of State Constant Combination

In considering the extensive application of the pseudo-critical point to the preparations of charts and tables for design purposes, it is of interest to examine the methods of estimating the values of T_{pc} and P_{pc} and improve the methods, if possible. Equations for the pseudo-critical can be derived by the methods used in combining equation of state constants. This is very complicated for equations of state containing more than two empirical constants. Joffe did this with the van der Waals equation to obtain two methods of estimating the pseudo-critical constants.

In developing both methods, Joffe combined the van der Waals a constant in the usual method, namely,

$$a^{1/2} = \Sigma x_i a_i^{1/2} \tag{195}$$

but he combined the b constant differently. In one method he used van der Waals' suggestion

$$b = \Sigma x_i b \tag{196}$$

while in the second he used the Lorentz, which is much more complex, being

$$b = 1/8 \sum_{ij} x_i x_j (b_i^{1/3} + b_j^{1/3})^3 \tag{197}$$

Using equations 195 and 196 along with equations 165

and 166 gave Joffe the following equations for the pseudo-critical point:

$$\frac{T_{pc}}{P_{pc}^{1/2}} = \sum x_i \frac{T_{c_i}}{P_{c_i}^{1/2}} \qquad (198)$$

$$\frac{T_{pc}}{P_{pc}} = \sum x_i \frac{T_{c_i}}{P_{c_i}} \qquad (199)$$

where: the terms are the same as defined under equations 193 and 194.

Equations 198 and 199 were shown to be slightly better than equations 193 and 194 but they are obviously more complex.

Joffe developed the other pair of equations by combining equations 195 and 197 with equations 165 and 166. The resulting equations from this combination have the objection of the complex Lorentz relationship. This second set of equations developed by Joffe are:

$$\frac{T_{pc}}{P_{pc}^{1/2}} = \sum x_i \frac{T_{c_i}}{P_{c_i}^{1/2}} \qquad (200)$$

$$\frac{T_{pc}}{P_{pc}} = 1/8 \sum_{i\,j} x_i x_j \left[\left(\frac{T_{c_i}}{P_{c_i}}\right)^{1/3} + \left(\frac{T_{c_j}}{P_{c_j}}\right)^{1/3} \right]^3 \qquad (201)$$

This last equation is evaluated by combining all components in the system in pairs, each one with every component present, including itself, in the way shown to the right of the summation sign and then summing up these binary combinations. This way of writing the equation, i.e., using "i" and "j," etc., is a mathematical shorthand way of expressing this combination.

From data available on pseudo-criticals, it cannot be concluded definitely that these more complex methods of finding the pseudo-criticals are justified. They are cartainly more complex and time consuming. Further investigations are necessary to settle this point.

Joffe compared these two methods, i.e. equations 198 and 199, and equations 200 and 201, with Kay's linear molal average method, equations 193 and 194 for five binary mixtures of N_2 CO_2 and hydrocarbons, and found that equations 200 and 201 method is the best of the three methods for a total of 305 P-V-T points, and that the method represented by equations 198 and 199 is only slightly better than the linear molal average method.

Following are the average and maximum deviations of computed volumes from observed volumes for the five mixtures by Kay's method and Joffe's second method.

	Kay Equations 193 and 194	Joffe—2nd Equations 200 and 201
Average Deviation, Percent	1.83 to 3.03	1.07 to 2.26
Maximum Deviation, Percent	3.75 to 10.2	3.17 to 4.59

This comparison is for binary mixtures. A more severe test on multicomponent mixtures, particularly where no one single component dominates the mixture, or is more than say 75 percent of the mixture, would be of interest.

Both of Joffe's methods were tried for the following four component hydrocarbon mixture and the resulting pseudo-critical temperatures and pressures

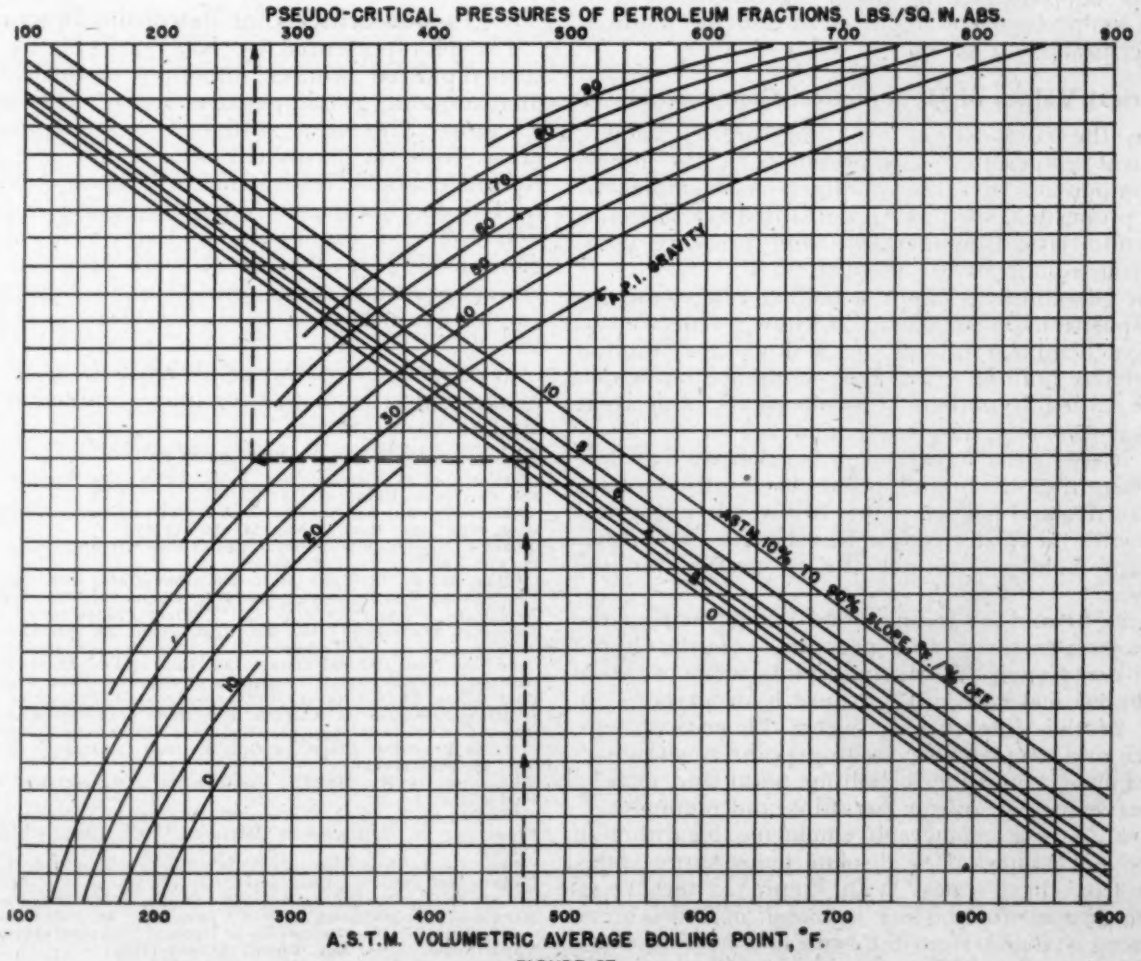


FIGURE 27

Pseudo-Critical Pressure as Function of V.A.B.P. and Gravity.

(Developed from W. B. Kay's correlation, Figure 23, by combining with molecular weight correlation, Figure 14 (Part VII), and definition of U.O.P. Characterization factor.)

were compared with the results obtained by Kay's method.

Composition	Mol. %
Methane	20
Ethane	20
Propane	30
Butane	30
Total	100

Pseudo-critical calculations by Kay's method (equations 193 and 194) and by Joffe's first (equations 198 and 199) and second (equations 200 and 201) methods are given below:

Pseudo-Critical	Kay's Method Equations 193 and 194	Joffe's 1st Method Equations 198 and 199	Joffe's 2nd Method Equations 200 and 201
T_{pc}, °R	608	616	626
P_{pc}, psia	627	618	640

Joffe's equations are too awkward for routine work. For the time being, it seems best to continue using Kay's method of computing pseudo-critical constants. If future P-V-T data on mixtures indicate that equations 193 and 194 are not adequate, it might be better to develop a residual correction term correlation for each equation than to derive entirely new relationships. This seems best because the simple linear molal average, represented by equations 193 and 194, is so close to the correct average that a residual would be of small magnitude. There is a possibility of using Joffe's equations to develop this graphical residual correction to the pseudo-critical temperatures and pressures. This work should be supplemented by more experimental P-V-T data on multicomponent mixtures to find the required pseudo-criticals.

Critical Values of Hypothetical Components

When the composition of a mixture is known, the critical temperatures and pressures for the individual components may be obtained from tables of critical properties, such as Table 13 in Part VII, and the pseudo-critical temperature and pressure computed from equations 193 and 194.

In the gasoline and naphtha boiling range, the exact compositions are not usually known. One way to handle such mixtures is to break down the fraction into narrow boiling groups of components, which may be called hypothetical components, and treat each as if it were a pure component.

This hypothetical component breakdown is made by dividing the true boiling point curve into plateaus. The gravities and boiling points of these hypothetical components make it possible to estimate their properties and combine to get the properties for the mixture.

Figures 24 and 25 are included for estimating the critical temperatures and pressures of the hypothetical components. Both these plots were prepared from the critical constants of pure hydrocarbons. In Figure 24 the difference between the critical temperature and atmospheric boiling point is shown as a function of atmospheric boiling point and gravity with curves drawn on for paraffins and olefins.

Figure 25 is a nomograph employing logarithm of pressure and reciprocal of absolute temperature scales, e.g. the Cox chart scales. With Figure 25 the critical pressure of a narrow boiling *hypothetical* hydrocarbon component is found from the critical temperature and atmospheric boiling point. An example illustrates the use of Figure 25 where a hypothetical component with a critical temperature of 510° F. and an atmospheric boiling temperature of 185° F., is found to have a critical pressure of 495 psia.

Critical temperatures and pressures from Figures 24 and 25 for hypothetical components are combined by equations 193 and 194 to give the linear molal average pseudo-critical conditions.

Pseudo-Criticals for Natural Gas

Brown and Holcomb[1] plotted the calculated (via Kay's method, equations 193 and 194) pseudo-critical temperatures and pressures against gas specific gravity (relative to air) for light hydrocarbon gaseous mixtures, i.e. natural gas between gravities of 0.55 and 0.9 (molecular weights of 16 to 26). The pseudo-critical pressure points fell on a straight line with good agreement. A straight line could be drawn through the pseudo-critical temperature points also although the agreement was not as good as for the pseudo-critical pressure.

Equations for this approximate correlation of Brown and Holcomb[1] (written by this author) are:

$$T_{pc} = 190 + 283G \quad (202)$$

$$P_{pc} = 710 - 60G \quad (203)$$

where: T_{pc} = pseudo-critical temperature, °R
P_{pc} = pseudo-critical pressure, psia.
G = gas gravity (air = 1.0)

$$G = \frac{\text{Mol. Wt.}}{29.0}$$

These equations are approximate and only for natural gas. It is preferable to compute the pseudo-critical temperatures and pressures from compositions and critical conditions for components where such data are available.

Pseudo-Criticals for Petroleum Fractions

The plot presented by Kay (see Figure 23) has been replotted as direct functions of the ASTM distillation and gravity properties of the petroleum fractions. This is a straight-forward operation because both the molecular weight and the Watson-Nelson characterization factor are functions of the ASTM distillation curve and the gravity. See equation 153 and Figure 14 of Part VII. The charts given as Figures 26 and 27 were prepared by combining equation 153, Figure 14 and Figure 23. Examples are indicated by dashed lines on Figures 26 and 27.

With pseudo-critical conditions, most of the thermodynamics developed for pure components may be applied to mixtures.

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(End of Part X. Part XI will appear in the May issue.)

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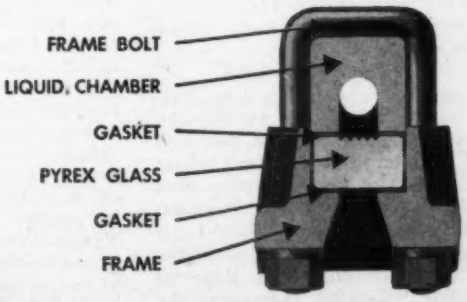

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EDWARD G. SCHEIBEL

A Lot Crowded Into His 31 Years

EDWARD G. SCHEIBEL, author of "Multi-Component Distillation Fractionation Calculations" (page 92) and "Nomograph for Equilibrium Flash Calculations" (soon to appear) is a comparative youngster as chemical engineers go, but he has a record that would do credit to one 15 years older. Catalogued it goes like this:

Born in Brooklyn in 1917.

Training—B.S. in Ch. E., Cooper Union Inst. of Tech., 1937; M. Ch. E., 1940 and D. Ch. E, 1943, Polytech. Inst. of Brooklyn; prof. eng. license, New York, 1944.

Positions—Plant supt. H. C. Bugbird Co., Inc., 1937-42 in charge of plant mfg. barium nitrate, potassium nitrate, strontium carbonate and strontium chloride.

Process eng., M. W. Kellogg Co., 1942-43 where he worked on design of light ends recovery plants, alkylation and isomerization units. (Biggest job—process design of extractive dist. units for separation of butane-butylene cut with acetone-water solvent.) Also worked on extractive dist. of toluene.

Process eng., Hydrocarbon Research, Inc., 1944-45 where he was primarily concerned with design of oxygen plants (with 8 patent applications taken out in his name on oxygen processes and equipment for the low temperature separation method.)

Currently connected with Hoffman-La Roche, Inc., where his main work is with development of various dist. and extraction processes but involves all fields of chem. eng.

Polytech. Inst. of Brooklyn, research associate, 1943-44; part-time instruction in Ch. E, since 1942; appointed adjunct prof. in Ch. E., 1945. Instituted and presents following courses: Calculation of absorption operations, calculation of extraction operations, calculation of dist. operations, interpretation of ch. eng. data, and instrumentation and process control.

Societies—Member ACS and Junior member A.I.Ch. E. and Chemists' Club of N. Y. Also member Phi Lambda Upsilon and Sigma Xi.

Other facts: Married, one child. Numerous hobbies including fishing, swimming bowling, pool, all card games, tennis, stamp and coin collector, photography. (Doesn't have a whole lot of time for hobbies, so he "concentrates on them one at a time.") Likes to do mechanical work, does all repairs on his car, also makes improvements on the Scheibel bungalow at Rockaway Point, Long Island. "Enjoys most just working around all day in my bathing suit and I look forward to spending summers at the ocean." Editor's Comment: Who wouldn't!)

Enjoys All Colorado University Victories

LYLE BRAY's contribution to this month's natural gasoline processing portfolio is entitled "Practical Considerations in Process Design of Fractionators" which commences on page 115.

Bray, who is a native Coloradan, was graduated from the state university at Boulder in 1941 with a B.S. in chemical engineering. Then shortly afterward, he went to work for Phillips Petroleum Company and he has retained that same connection throughout his professional career.

His first work for Phillips was in that company's natural gas and gasoline department. In 1942-43 he worked on the process design of fractionators for the Plains butadiene plant, located at Borger, Texas. After that plant started operation in the summer of 1943, he served as a process engineer at the plant, again in the fractionation section, until the fall of 1946, during the last year of which he was section leader. Since that time he has been a technical assistant in the Bartlesville, Okla., office of the Defense Plant Division, which operates the Plains plant.

So much for his working days.

Bray is married and has a son, Patrick, who is 2½ years old. In Borger, he served as a scoutmaster of the Boy Scouts and although Dad doesn't admit it, Patrick must be taking more time because Bray hasn't been active in scouting since moving to Bartlesville.

Most sports, particularly those of the University of Colorado, claim Bray as an enthusiastic follower. However, the only sport in which he is active is golf. His score? Well, if you can't break 80 pretty regularly, then you won't have to give him any strokes.

Bray is a member of the First Presbyterian church, the Junior Chamber of Commerce, American Institute of Chemical Engineers, Tau Beta Pi, Sigma Tau and Alpha Chi Sigma.

JOSEPH E. BLUDWORTH

Betting on Future Of Chemical Industry

JOSEPH E. Bludworth, author of the article beginning on page 110 of this issue, "Some Aspects of the Industrial Oxidation of Lower Aliphatic Hydrocarbons," is a petroleum and chemical consulting engineer with offices in Corpus Christi and Houston, Texas. He has been active in the design and construction of various petroleum chemical plants in the Gulf Coast area. Currently, he is coordinating the chemical engineering for the $15 million McCarthy Chemical Company plant at Winnie, Texas, with Fomac Engineers, the Houston firm with which he is associated.

Bludworth hails (believe it or not) from Flatonia, Texas (about 90 miles from San Antonio), and entered the petroleum industry by working for the old firm of Chestnut and Smith (now part of Lone Star Gas Company), where he assisted other company technicians in development work on high pressure natural gas processing.

A hard working, straight-forward individual, Bludworth decided to stake his future on his confidence that the chemical industry was in for big things. Later he opened consulting offices in Dallas, Texas. While there he perfected several processes that attracted national interest in the chemical industry.

In 1934 he joined Celanese Corporation of America, where he was director of petroleum chemicals research and development until he returned to consulting work some 12 years later. He moved to Corpus Christi in 1943, when ground was broken for the Celanese Chemical plant at Bishop, which was designed and constructed under his supervision. This plant reputedly produces more formalde-

hyde (one of the plant's many chemicals) than was produced in the U. S. before World War II.

Bludworth believes that the chemical plants now being built are only the forerunners of a rapidly expanding new phase of industry involving mass production by simple methods of chemicals from petroleum, particularly natural gas. In his opinion, the chemist can ultimately achieve any desired results with a hydrocarbon, and he has pointed out that all hydrocarbons, including those considered waste products, are suitable for making valuable compounds. He insists, however, that the objective be based on market demand before development is continued.

Colwyn Medal Goes To Dinsmore in June

In recognition of his work in synthetic rubber research, development and application to product, and as a tribute to the American rubber industry's achievements in this field during the war, presentation of the Colwyn Gold Medal for 1947 will be made June 25 in London, England, by the Institution of the Rubber Industry to Dr. Ray P. Dinsmore, vice president in charge of research for The Goodyear Tire & Rubber Company.

Dr. Dinsmore, who was assistant deputy rubber director for the United States during the war until May, 1943, organized and directed research and development work on synthetic rubber for the government and coordinated the activities of both industrial and university laboratories, after which he returned to industry to follow through on synthetic rubber's application to rubber product manufacture.

The Colwyn Medal is awarded for conspicuous services of a scientific or technical character, having an important bearing upon the rubber industry.

California Standard Position for Brown

G. Stewart Brown, former State Department official, has been named manager of the public relations department of Standard Oil Company of California.

Brown, for many years a foreign correspondent in Europe, has had a wide variety of experience, including service with the American Red Cross as national vice chairman in charge of public relations and as director of public relations of the Allied Commission in Italy.

In 1946, when he returned to Washington, he was appointed deputy director of the Office of Information and Educational Exchange, which position he resigned to become associated with Standard.

Scholarship Award

Ruffin G. Stirling, Jr., son of R. G. Stirling of the mechanical and construction division office, Standard Oil Company of New Jersey, Baton Rouge, La., has been awarded a Teagle Foundation scholarship for advanced study in chemical engineering and on February 9 entered Massachusetts Institute of Technology.

Young Stirling is the fourth from Baton Rouge to receive one of the awards, established in 1945 by Walter C. Teagle, former president of Standard Oil Company (New Jersey).

DR. ROBERT S. ARIES

Aries Paper Is Another Contribution From Brooklyn Poly

IT APPEARS that Polytechnic Institute of Brooklyn is making a major contribution to this month's PETROLEUM REFINER.

In a previous column there are references to Edward Scheibel's article and the fact that Mr. Scheibel is engaged on a part-time basis teaching at that school. Here in this space data about another author, similarly engaged, appear.

Dr. Robert S. Aries, whose "Synthetic Alcohol from Petroleum" discussion starts on page 124, is an adjunct professor of chemical engineering of Brooklyn Polytechnic, where he does development work on petroleum chemicals and teaches in the graduate school.

Heads Own Firm

He also heads the firm of R. S. Aries and Associates, consulting engineers and economists and is engaged in consultation work in this country, England and France. His interests include the manufacturing and uses of alcohol, ethylene oxide and glycol, and the separation of oxygenated products from the Fischer-Tropsch synthesis. He has authored more than three score articles of a technical and economic nature on the chemical process industries.

For all this work Dr. Aries was well trained. He attended the Universities of Minnesota, Polytechnic Institute of Brooklyn, Yale and Columbia. In addition to his three degrees in engineering, he also has a master's degree in business administration and, as he says, has "always tried to link chemical engineering and economics in my career." This is well exemplified by the process and cost evaluation in his current piece.

Likes Foreign Travel

In the time which Dr. Aries has aside from his professional activities, he manages to do a little skiing and also play tennis. He travels in foreign lands whenever possible and his hope is that there will come a time when all tariff barriers will be broken down so that each country can make and export whatever each is best suited for.

Melamid Resumes Belgian Refinery Post

Dr. Michael Melamid, active in New York with the Melanol Corporation, has resumed his former position as technical director of the Raffinerie Belge de Petrole at Anvers. This is one of the two European plants whereby Dr. Melamid's process Mexical Panuco oil is refined for the recovery of asphalt. The other plant is the Compagnie des Produits Chimiques et Raffineries de Berre, near Marseilles. During the war the Belgian plant ceased operations but is now again working at full capacity (150,000 tons yearly).

Dr. Melamid is at present occupied with adapting his process to the refining of Arabian oil.

Two New Members Of Information Group

Two new members of the Oil Industry Information Committee, announced last month are, Stanton Smith, president of Smith Oil and Refining Company, and G. Stewart Brown, manager of the public relations department of Standard Oil Company of California.

Brown recently was appointed to his position with Standard, having formerly been with the State Department. Smith is chairman of the Illinois Oil Industry Information Committee and a former director of the Illinois Petroleum Marketers Association.

Instrument Society Press Head Chosen

Hank Cabot, director of public relations for the Brown Instrument division of Minneapolis-Honeywell Regulator Company, has been appointed chairman of the press relations committee for the 1948 national conference and exhibit of the Instrument Society of America, E. J. Grace, chief instrument engineer of Sun Oil Company and chairman of the ISA meetings committee, has announced.

The 1948 conference and exhibit will be held at Convention Hall, Philadelphia, September 13 to 17.

Refinery and Other Plant Construction

Companies whose projects are not listed or readers who know of unlisted projects are urged to: 1) send in data on units not listed; 2) furnish details which are missing in the current tabulation and 3) report from time to time on progress of such work. Such cooperation will be very much appreciated.

COMPANY	Plant Site	Project	Daily Capacity	Estimated Cost	Status	Probable Completion	Licensor	Engineering	Contractor
EAST COAST									
Aetna Oil Co.	Louisville, Ky.	Gasoline Recovery Unit	800 mcf	$75,000	Under Constr.	.....	.....	Aetna	Aetna
Atlantic Ref. Co.	Philadelphia, Pa.	Propane, lubes Wax Plant	3,800 bbls.	.....	Under Constr.	Early '48	JUIK, Texaco	Badger, Atlantic Kellogg & Lummus	Badger, Lummus, Kellogg, Atlantic, Lauter Constr.
Gulf Oil Corp.	Philadelphia, Pa.	Naptha, Polyform	16,000 bbls. (circ.)	.....	Under Constr.	.....	Gulf-Phillips	Lummus	Lummus
*Gulf Oil Corp.	Philadelphia, Pa.	Vacuum Unit	12,000 bbls.	.....	Under Constr.	.....	.....	Lummus	Lummus
Socony-Vacuum Oil Co., Inc.	Paulsboro, N.J.	Crude, Lube Plant	.....	Part of $45 million	Under Constr.	Jan. 1, 1949	Texaco	Badger	Badger
Standard Oil Co. (New Jersey)	Linden, New Jersey	Vac. Cat Cracking, pipe still coker	90,000 bbls.	.....	Contract let	1st Qtr. 1949	.....	S.I.D. & Braun	Braun
Sun Oil Co.	Marcus Hook, Pa.	Lube, Wax, Anti-Pollution Project	.....	.....	.....	.....	.....	.....	.....
The Texas Co.	Eagle Pt., New Jersey	Vac. still, Cat Cracking	40,000 bbls.	.....	Contract let	Late '49	.....	Kellogg, Texaco Co.	Kellogg, Foster-Wheeler
MID-CONTINENT									
Barsco Oil Co.	Wichita, Kans.	Cat. desul. Reforming	4,000 bbls.	$500,000	Under Constr.	June, '48	Phillips	Koch Eng. Co.	Koch Eng.
Berry Asphalt Co.	Waterloo, Ark.	Lub Unit, Expand Asphalt facilities	1,200 bbls.	.....	.....	Mid '48	.....	.....	.....
Berry Asphalt Co.	Stephens, Ark.	Expand Asphalt Facilities	2,000 bbls.	.....	.....	Mid '48	.....	.....	.....
Cities Service Oil Co.	Chicago	Compounding & Grease Plant	1,500 bbls.	$2.5 million	Design	Early '49	.....	Sumner-Sollitt	.....
Cities Service Oil Co. (Delaware)	E. Chicago, Ill	Modernize plant	.....	$18 million	Under Constr.	Late '48	Kellogg	Kellogg	Kellogg
Cities Service Oil Co. (Delaware)	E. Chicago, Ill	Laboratory	.....	$300,000	Under Constr.	Spring '48	.....	Cities Service	Sumner-Sollitt
Consumer Co-operative Association	Coffeyville, Kansas	MEK Dewaxing	2500 bbls.	$1.5 million	Contracted	1949	Texaco	Process Eng.	Process Eng.
Fen Tex. Ref. Co.[6]	Wynnewood, Okla.	Asphalt Plant	.....	.....	.....	.....	.....	.....	.....
Gulf Rfg. Co.	Toledo, O.	Top & Vac. Unit	32,000 bbls.	.....	Under Constr.	Oct., 1948	.....	Badger	Badger
*Gulf Rfg. Co.	Toledo, O.	Fluid Cat. Crk.	15,500 bbls.	.....	Under Constr.	May, 1949	.....	.....	Kellogg
*Gulf Rfg. Co.	Toledo, O.	Polyform Unit	5,000 bbls.	.....	Under Constr.	May, 1949	U.O.P.	.....	Kellogg
*Gulf Ref. Co.	Cincinnati	Fluid Unit	15,500 bbls.	.....	Under Constr.	June, 1949	.....	.....	Kellogg
*Gulf Rfg. Co.	Cincinnati	Polyform Plant	5,000 bbls.	.....	Under Constr.	June, 1949	U.O.P.	.....	Kellogg
Otha A. Grimes	Holdenville, Okla.	Gasoline Plant	4[illegible]000 mcf	.....	.....	.....	.....	.....	.....
Hagy, Harrington & Marsh	Hooker, Okla.	Gasoline and LPG Recovery	100,000 mcf	$2 million	Under Constr.	July 1, '48	.....	Pritchard	Pritchard
Hiwan Oil Co.	Magnolia, Ark.	Gasoline Plant	.....	.....	First Stage Completed	.....	.....	Petro. Eng.	Petro. Eng.
Kanotex Refg. Co.	Arkansas City, Kans.	Cat Cracking	4,200 bbls.	$500,000	Contract let	Early '48	Phillips	Koch Eng.	Koch Eng.
Mid-Continent Pet. Co.	Tulsa, Okla.	Vac still, Cat Cracking	45,000 bbls.	$8.5 million	Contract let	June, '48	Kellogg	Kellogg	Kellogg
*Midland Cooperative Wholesale Co.	Cushing, Okla.	Thermal Reforming	1,400 bbls.	Part of $500,000	Under Constr.	Aug., 1948	U.O.P.	U.O.P. and Ref. Eng. Co.	Ref. Eng. Co.
*Midland Cooperative Wholesale Co.	Cushing, Okla.	Increase Topping Capacity	6,500 bbls.	Part of $500,000	Under Constr.	Aug., 1948	U.O.P.	U.O.P. and Ref. Eng. Co.	Ref. Eng. Co.
Ohio Oil Co.	Robinson, Ill.	Vac Unit, Cat Cracking	27,500 bbls.	$15 million	Under Constr.	Late '48	U.O.P.	McKee	McKee, Pritchard
Old Dutch Ref. Co.	Muskegon, Mich.	LPG Recovery	3,400 gallons	$45,000	.....	.....	U.O.P.	Old Dutch	Old Dutch
Petco Corporation	Blue Island, Ill.	Skim. & Crkg.	10,000 bbls.	.....	Under Constr.	Early '48	.....	Clark's	Clark's
Phillips Pet. Co.	Kansas City, Kansas	Cat Cracking, Poly and Vacuum Units	.....	.....	Contracted	.....	.....	Braun	Braun
Rock Island Rfg. Corp.	Rock Island, Ind.	Fluid Cat. Crack.	10,000 bbls.	$1,200,000	Contracted	Early '49	U.O.P.	U.O.P.	Own Staff
Root Pet. Co.	El Dorado, Ark.	Revision of Cat. Cracker	7,000 bbls.	$125,000	Under Constr.	May, 1948	.....	U.O.P.	Mason Bros.
Sinclair Refg. Co.	East Chicago, Ill.	Crude Cat. Cracker, Poly Plant & Lube Treating	55,000 bbls.	.....	Under Constr.	Early '48	.....	Lummus	Braun, Lummus
Skelly Oil Co.	Stephens Co., Okla.	Gasoline plant	.....	.....	.....	.....	.....	.....	Born
Socony-Vacuum Oil Co., Inc.	E. St. Louis, Ill.	Thermal Cracker & distillation	.....	Part of $45 million	Contract let	Jan. 1, 1949	.....	Bechtel Corp.	Bechtel Corp.
Socony-Vacuum Oil Co., Inc.	Augusta, Kans.	Cat Cracking	10,000 bbls.	Part of $45 million	Projected	.....	.....	.....	.....
Stanolind O & G Co.	Hugoton Field, Kansas	Gasoline Plant	150,000 mcf	.....	Approved	.....	.....	J & L Constr.	J & L Constr.
Stanolind O & G Co.	Garden City, Kas.	Gas Syn. Plant	100,000 mcf	$80 million	Constr. in Fall, '48	Jan., 1950	.....	Hydro Research	Kelogg, Foster-Wheeler, Stone & Webster
Stanolind O & G Co.	Hugoton Field, Kansas	Chemical Plant	130,000 lbs./yr.	.....	Announced	.....	.....	Badger	Badger
Stanolind Oil & Gas Co.	Ulysses, Kan.	Gasoline Plant	100,000 mcf.	.....	Projected	.....	.....	.....	.....
*Standard Oil Co. (Indiana)	Casper, Wyo.	Cat Crack.	8,000 bbls.	.....	Under Constr.	Early 1949	.....	.....	Kellogg
Standard Oil Co. (Indiana)	Sugar Creek, Missouri	Cat Cracking	25,000 bbls.	.....	.....	Spring '48	.....	.....	Kellogg
Standard Oil Co. (Indiana)	Whiting, Indiana	Cat Cracking	25,000 bbls.	.....	Under Constr.	Spring '48	.....	Lummus S.O.D. & Braun	Kellogg
*Standard Oil Co. (Indiana)	Whiting, Ind.	Motor Oils Unit	.....	.....	Under Constr.	Mid 1948	.....	.....	Foster-Wheeler, Stone & Webster Kellogg
Standard Oil Co. (Ohio)	Lima, Ohio	Cat Cracking	16,000 bbls.	$10 million	Contract let	Late '49	U.O.P.	McKee	McKee

■ CONTINUED ON PAGE 152

Another Example of Efficient Power at Lower Cost

Why the new Turboflow development is of major importance!

1. IT GUARANTEES 15% LESS FUEL CONSUMPTION!

The Turboflow engine development offers fuel economy that cannot be approached by any other 2-cycle gas engine. Instead of the customary guarantee of 10,000 Btu consumption per hp hour, the Turboflow *guarantee* is actually 8,500 Btu's per hp hour—a full 15% saving in fuel! This remarkable performance has been thoroughly proved during the past two years. It is readily obtained with any of the Cooper-Bessemer Turboflow V-angle types or sizes. *The Turboflow principle should not be confused with supercharging.* It introduces ignition refinements and *internal* design changes which result in a much more intimate mixture of air and gas for more efficient combustion and superior heat utilization.

2. IT INCREASES THE POWER-TO-SPACE RATIO A FULL 10%.

Pulling at least 10% more maximum load than the same size old style engine, the Turboflow engine easily carries 10% more continuous load without encroaching on its overload capacity. Moreover, exhaust temperature is actually *less* than that of the old style engine. Obviously, a 10% higher hp rating is completely justified . . . *and without any increase whatever in foundation, housing or piping requirements!*

3. IT AFFORDS GREAT INSTALLATION AND OPERATING SAVINGS!

Entirely aside from its remarkable fuel economy, the Turboflow development obviously offers other great savings. With its greater power-to-space ratio, the over-all cost of installation per horsepower is far less! Since it takes fewer engines and actually *less* space to meet total power needs, the cost per horsepower of personnel, maintenance, part replacement and general overhead is correspondingly lowered! Furthermore, the Turboflow principle *reduces heat dissipation both to water and lube oil*, so that cooling towers, radiators or heat exchangers need have no greater capacity than for old style, less powerful engines.

Yes, no matter how you look at it, the Cooper-Bessemer Turboflow development offers tremendous savings. Why not check with the nearest Cooper-Bessemer office on the adaptability of this great development to your particular needs?

by Cooper-Bessemer

MOUNT VERNON, OHIO AND GROVE CITY, PENNA.

uston, Dallas, Greggton, Pampa and Odessa, Texas Tulsa Shreveport St. Louis Los Angeles

REFINERY AND OTHER PLANT CONSTRUCTION (Continued)

COMPANY	Plant Site	Project	Daily Capacity	Estimated Cost	Status	Probable Completion	Licensor	Engineering	Contractor
Sun Oil Co.	Toledo, Ohio	Anti-Pollution Project	.....	.....	.....	.....	.....	.....	.....
Sunray Oil Co.	Duncan, Okla.	Crude Plant	20,000 bbls.	.....	Under Constr.	April, 1948	.....	U.O.P.	Ref. Eng. Co.
*Sunray Oil Corp.	Duncan, Okla.	Cat. Poly.	1000 bbls.	$271,000	Under Constr.	May, 1948	U.O.P.	U.O.P. and Ref. Eng. Co.	Ref. Eng. Co.
*Sunray Oil Corp.	Duncan, Okla.	2 Treating Units	.....	.....	Under Constr.	.....	U.O.P.	Ref. Eng. Co.	Ref. Eng. Co.
Sunray Oil Corp.	Duncan, Okla.	Revision Cat. Cracker	10,000 bbls.	$150,000	Under Constr.	May, 1948	U.O.P.	U.O.P.	Mason Bros.
The Texas Co.	Lawrenceville, Illinois	Crude and Cat Cracking	20,000 bbls.	.....	Announced	1950	.....	Foster-Wheeler	Kellogg, Foster-Wheeler
The Texas Co.	W. Tulsa, Okla.	Crude and Cat Cracking	30,000 bbls.	.....	Contract let	1949	.....	Texas Co., Kellogg, Foster-Wheeler	Kellogg, Foster-Wheeler
*Trans-Marine O. Rfg. Co.	S. Milwaukee, Wis.	Crude, Vac., Refrm, Desulf.	5,000 bbls.	$1.3 million	Under Constr.	.....	.....	.....	Ref. Eng. Co.
Warren Pet. Corp.	Madill, Okla.	Gaso. Plant Enlargement	1,500 bbls.	.....	Projected	Jan. 1, 1949	.....	Hudson	Hudson
Warren Pet. Corp.	Maysville, Okla.	Gasoline Plant	8,000 bbls.	.....	Under Constr.	Jan. 1, 1949	.....	Dresser	Dresser
Wood River Oil & Refg. Co.	Hartford, Illinois	Cat Cracking, gasoil, Cracking	8,500 bbls.	$1 million	Under Constr.	Late '48	Phillips	Koch Eng.	Koch Eng.
SOUTHWEST									
American Republics Corp.	Silsbee, Texas	Gasoline Plant	.....	.....	.....	.....	.....	.....	Gasoline Plant Constr. Co.
*Arkla Oil Co.	Rodessa, La.	Add Compressors	60,000 mcf	$325,000	.....	.....	.....	.....	Own Staff
Atlantic Refg. Co.	Dallas, Texas	Geop. Lab. Bldg.	.....	.....	Under Constr.	'48	.....	Atlantic	Wohlfeld
Barnsdall Oil Co.	Benton, La.	Gasoline Plant	.....	.....	Under Constr.	.....	.....	.....	Lummus
Calumet Refg. Co.	Princeton, La.	Modernization of Lube Plant	700 bbls.	.....	Under Constr.	Early '49	.....	.....	.....
Carthage Corp.	Carthage, Field	Gasoline Plant	100,000 mcf	$3.5 million	Under Constr.	Jan. 1, '49	.....	Hudson	Hudson
Carthage Hydrocol, Inc.	Brownsville, Tex.	Hydrocol	64,000 mcf	$19 million	Under Constr.	1948	Hydro-Res. Inc., Texaco	McKee	McKee
*The Chicago Corp.	Carthage, Tex.	Enlarge Gas. Plt.	210,000 mcf	$2 million	Authorized	June, 1949	.....	Own Staff	.....
Citcon Corp.[1]	L. Charles, La.	Lube Plant	6,000 bbls.	$30 million	Under Constr.	.....	Texaco, Max B. Miller	Max B. Miller Lummus Citcon	Lummus, Max B. Miller
Continental Oil Co.	Sonora, Texas	Gasoline Plant	.....	.....	.....	.....	.....	.....	Brown & Root
Ethyl Corp.	Baton Rouge, La.	Mfg. Facilities	.....	$20 million	Announcement	.....	.....	.....	.....
Gray-Wolfe Co.	Montgomery Co.	Gasoline Plant	13,000 mcf	$2 million	Under Constr.	Jan., 1949	.....	O. L. Olsen	O. L. Olsen
Gulf Oil Corp.	Crane Co., Tex.	Gasoline Plant	23,000 mcf	.....	.....	.....	.....	.....	.....
Gulf Oil Corp.	Pt. Arthur, Tex.	3 Top. & Vac. units	22,000 bbls. ea.	.....	Under Constr.	April-June, 1948	.....	.....	Badger
Gulf Oil Corp.	Pt. Arthur, Tex.	Cat. desulf.	17,000 bbls.	.....	Under Constr.	.....	.....	Lummus	Lummus
Gulf Oil Corp.	Pt. Arthur, Tex.	Naphtha Polyform.	32,000 bbls. circ.	.....	Under Constr.	.....	.....	Lummus	Lummus
Humble O & R Co.	East Texas	Gasoline Plant	.....	$2 million	Under Constr.	.....	.....	Fish Eng. Corp.	.....
Humble O & R Co.	Katy, Texas	Propane, Butane Extraction	19,000 bbls.	$7 million	Contract let	.....	.....	.....	Stearns-Roger
Humble O & R Co.	Refugio, Texas	Gasoline Plant	35,000 mcf	$2.5 million	Under Constr.	1948	.....	.....	Gaso. Plant
Humble O & R Co.	Baytown, Texas	Lube dewaxing	4,600 bbls.	$5 million	Under Constr.	Early '48	Texaco	Badger	Badger
Humble O & R Co.	Kilgore, Texas	Gasoline Plant	150,000 gallons	$2 million	Under Constr.	.....	.....	Fish	Fish
Jefferson Chem. Co	Pt. Neches, Tex.	Petro-chem	.....	.....	Under Constr.	1948	.....	Jefferson	Jefferson, Badger, Lummus
Magnolia Pet. Co.	Beaumont, Tex.	Lube Plant	.....	.....	Under Constr.	Jan. 1, 1949	Texaco	Lummus	Lummus, Kellogg
Magnolia Pet. Co.	Seeligson Field, Texas	Gasoline Plant	90,000 mcf	.....	.....	Late '48	.....	Hudson	.....
Magnolia Pet. Co.	Mamou, La.	Gasoline Plant	7,000 mcf	.....	Under Constr.	Late '48	.....	.....	Gaso. Plant
Fred M. Manning, Inc.	Stephens Co., Texas	Pressure Maintenance	20,000 mcf	$500,000	Contract let	Completed, testing	.....	Fish Eng. Corp.	.....
McCarthy Chem. Co.	Winnie, Texas	Petro-chem, Cat Cracking	5,000 bbls.	$3 million	Under Constr.	.....	Phillips	Fish Eng. Corp.	.....
Monsanto Chem. Co.	Texas City, Tex.	Rebuilding Petro-chem	.....	$6 million	Contract let	.....	.....	.....	Bellows Co., Leon Con. Co.
Peteo Corporation	Marrero, La.	Skim. & Crkg.	5,000 bbls.	.....	Under Constr.	Early '48	.....	Clark's	Clark's
Petrol Terminals	Texas City, Tex.	Cat Cracker (recondition)	.....	.....	Under Constr.	.....	.....	.....	Cat. Constr.
Phillips Pet. Co.	Hansford Co., Texas	Gasoline & LPG Plant	343,000 mcf	.....	Announced	.....	.....	Own Staff	.....
Plomo Co.	Agua Dulce Feld	Processing Plant	30,000 mcf	$1 million	Under Constr.	"Soon"	.....	.....	.....
Pure Oil Co.	Smith's Bluff, Texas	Lube Plant	.....	$9 million	Contract let	.....	Texaco	Lummus	Lummus
*Republic O. Rfg. Co.	Texas City, Tex.	Plant rehabilitation	30,000 bbls.	$500,000	Making Survey	.....	U.O.P.	U.O.P.	.....
Richardson and Base	Keystone Field, Texas	Gasoline Plant	70,000 mcf	.....	Constr. Appr'vd	10-1-48	.....	J & L Constr.	J & L Constr.
Shell Chemical Corp.	Houston, Texas	Glycerine	.....	Part $35 mln.	Under Constr.	Early '48	.....	.....	Braun
Shell Oil Co.	Notrees, Texas	Gasoline Plant	45,000 mcf	$4 million	Under Constr.	April, '48	.....	Brown & Root	Brown & Root
Shell Oil Co.	Houston, Texas	Lube Unit	2,800 bbls.	$10 million	Under Constr.	July '48	Texaco	Lummus	Lummus, Kellogg, Fluor
Shell Oil Co.	Denver City, Tex.	Gasoline Plant	24,000 mcf	.....	Under Constr.	1948	.....	.....	Fluor Corp.
Shell Oil Co.	Norco, La.	Thermal Cracking, Asphalt Plt.	12,000 bbls.	.....	.....	April '48	.....	Shell	Shell
Sid Richardson Carbon Co.	Odessa, Texas	Carbon black Plant	194,500 lbs.	.....	Under Constr.	.....	.....	.....	.....
Sid Richardson Gasoline Co	Kermit, Texas	Gasoline & LPG Plant	200,000 gallons	.....	.....	.....	.....	.....	.....
Sinclair Refg. Co.	Houston, Texas	Crude Still, Barrel House, Grease Works.	30,000 bbls.	.....	Under Constr.	.....	.....	.....	Fluor
Southwestern Oil & Refg. Co.	Corpus Christi, Texas	Catalytic Reforming	5,000 bbls.	$800,000	Under Constr.	August, '48	.....	Process Eng.	Process Eng.
Stanolind O & G Co.	Slaughter, Texas	Gasoline Plant	90,000 mcf	.....	Under Constr.	Dec., 1948	.....	Braun	Braun
Stanolind O & G Co.	Hockley County, Texas	Gasoline Plant	.....	.....	.....	.....	.....	.....	.....
Stanolind O & G Co.	Brownsville, Tex.	Chemical Plant	130,000 lbs./yr.	.....	Announced	.....	.....	Badger	Badger
Stanolind O & G Co.	Pettus, Texas	Cycling Plant	159,000 mcf	.....	Projected	1949	.....	J & L Constr.	J & L Constr.
Standard Oil Co. (New Jersey)	Baton Rouge, Louisiana	Lub. Vac. still, Crude, Dewax, poly.	111,500 bbls.	.....	Under Constr.	Early '48	Texaco	Badger, S.O.D., Braun, McKee, Kellogg	Badger, Kellogg, McKee, Braun
Sun Oil Co.	Richland Parish, Louisiana	Gasoline Plant	70,000 gal.	.....	Under Constr.	.....	.....	.....	Pet. Eng., Inc.
*Sun Oil Co.	Starr Co., Tex.	Gasoline Plant	25,000 mcf	Part of $12 mil.	Bids asked	April, 1949	.....	.....	.....
Talco Asphalt & Refining Division American Liberty Oil Co.	Mt. Pleasant, Texas	Poly L.P.G. Plant	26,800 mcf	$225,000	Under Constr.	April 15, '48	U.O.P.	Talco	Talco
The Texas Co.	Port Arthur, Texas	Lube & Wax additions	2,500 bbls. oil	.....	Under Constr.	Late '48	.....	Texas Co., Foster-Wheeler	Foster-Wheeler

REFINERY AND OTHER PLANT CONSTRUCTION (Continued)

COMPANY	Plant Site	Project	Daily Capacity	Estimated Cost	Status	Probable Completion	Licensor	Engineering	Contractor
*The Texas Co.	Lafitte, La.	Expand Gaso-Plant	6,500 mcf	.....	.....	.....	.....	Hudson	Hudson
*United Gas Co.	Carthage, Tex.	2nd Gas. Plant	100,000 mcf	$3.5 million	Announced	Feb., 1949	.....	Hudson	Hudson
*Waggoner Est. Ref.	Electra, Tex.	Crack Unit Revisions	.....	$250,000	Under Constr.	Fall, 1948	.....	Ref. Eng. Co.	Ref. Eng. Co.
Warren Pet. Corp.	Breckenridge, Texas	L.P.G. Extraction	1,500 bbls.	.....	Under Constr.	Jan. 1, 1949	.....	Warren	Warren
Warren Pet. Corp.	Holliday, Tex.	Expansion & Propane prod.	2,500 bbls.	.....	Under Constr.	Jan. 1, 1949	.....	Warren	Warren
Winnsboro Refg. Co.	Winnsboro, Tex.	Crude Plant	3,000 bbls.	$175,000	Under Constr.	Early '48	.....	.....	Mid-Continent
Witco Hydrocarbon Corp.	Big Lake, Texas	Gasoline Plant	2,500 bbls.	$2.6 million	Under Constr.	July '48	.....	Witco	Fish
Witco Hydrocarbon Corp.	Witco, Texas	Carbon Black	.....	Part of $3.5 million	Contracted	.....	.....	Fish	Fish & Witco
ROCKY MOUNTAIN									
Bureau of Mines	Rifle, Colo.	Shale Oil Refy.	3000 bbls.	$450,000	Under Constr.	Fall, 1948	.....	Ref. Eng. Co.	Ref. Eng. Co.
The California Oil Company	Rangely, Colo.	Gasoline Plant	25,000 gals.	$3 million	Under Constr.	'48	.....	Hudson	Hudson
Carter Oil Co.	Billings, Mont.	Crude Cat Crk.	30,000 bbls.	$20 million	Under Constr.	3rd Qtr. 1949	.....	U.O.P., S.O.D.	Fluor Corp.
Continental Oil Co.	Billings, Mont.	Cat Cracking	7,500 bbls.	$8.5 million	Contract signed	1949	U.O.P.	U.O.P.	J & L Constr.
Continental Oil Co.	Denver, Colo.	Crude, Cat Crk.	7,500 bbls.	$4 million	Under Constr.	Fall 1948	Houdry Corp. & U.O.P.	Con. Oil Co., Lummus	Lummus
Gulf Oil Corp.	Eunice, N. M.	Gasoline Plant	.....	.....	.....	.....	.....	Hudson	Hudson
Skelly Oil Co.	Lea County, N. M.	Expanding Gasoline plant	.....	.....	.....	.....	.....	.....	J. E. Carlson
Socony-Vacuum Oil Co., Inc.	Casper, Wyo.	Cat Cracking	4,000 bbls.	Part of $45 million	Under Constr.	Jan. 1, 1949	.....	Cat. Constr.	Cat. Constr.
Stanolind O & G Co.	Elk Basin, Wyo.	Gasoline Plant	12,000 mcf	.....	Approved	.....	.....	Stone & Webster	Stone & Webster
Standard Oil Co. of Calif.	Salt Lake City, Utah	Crude dist. unit	25,000 bbls.	$5 million	Under Constr.	Nov., 1948	.....	.....	Bechtel
Utah Oil Refg. Co.	Salt Lake City, Utah	Expand existing Refinery	25,000 bbls.	$2.5 million	Announced	Late '48	.....	.....	.....
Warren Pet. Corp.	Monument, N.M.	Gasoline Plant Enlargement	2,500 bbls.	.....	Under Constr.	Jan 1, 1949	.....	Warren	Dresser Eng.
WEST COAST									
General Pet. Corp.	Torrance, Calif.	Vacuum flash tower	32,000 bbls.	$558,000	Contracted	3rd Qtr. '48	.....	Braun	Braun
Hancock Chem. Co.	Watson, Calif.	Elemental Sulfur Plant	50 tons	$1 million	Engineering being completed	Nov., 1948	.....	Badger	Badger
*Shell Chemical Co.	Martinez, Calif.	Pet. Additives Plant	.....	.....	Under Constr.	Late 1948	.....	Parsons	Parsons
*Robt. S. Lytle, Opr.	Coalinga, Calif.	H_2S Removal Plant	33,000 mcf	$175,000	Engineering in Course	1 year	Fluor	Fluor	Fluor
Shell Chemical Corp.	Shell Pt. Calif.	Syn. Ammonia	.....	.....	.....	Early '48	.....	.....	.....
Standard Oil Co., of Calif.	Bakersfield, Calif.	Crude dist, Cat Cracking	10,000 bbls.	$7.5 million	Announced	Mid '49	Braun	Braun	Braun
*Standard Oil of Calif.	Bakersfield, Calif.	Thermal Crk., Gaso. Stabil., Gaso. Treating	9,000 bbls.	.....	Contracted	.....	U.O.P.	U.O.P.	Braun
Standard Oil Co. of Calif.	El Segundo, Cal.	Filling Plant Modernization	.....	$2 million	Under Constr.	Dec. '48	.....	.....	.....
Standard Oil Co. of Calif.	Richmond, (California)	Grease, Wax Plant, Dock Facilities	82,000 lbs.	$13 million	Under Constr.	Sept. '48	U.O.P.	W. G. Peugh	Barrett & Hilp, Swinerton & Walberg & Bon C. Gerwick Inc., Humiston-Rosendahl Inc.
FOREIGN									
Agwi Pet. Corp.	Fawley, Eng.	.....	120,000 bbls.	$140 million	Projected	1952	.....	.....	.....
British Pet. Chem., Ltd.	Grangemouth, Scotland	Petro-chem	.....	.....	.....	.....	.....	.....	.....
Caltex Pet. Maatschappij, N.V.	Rotterdam, Holland	Crude Plant	20,000 bbls.	$18 million	Projected	Early '50	.....	McKee	McKee
Consolidated Ref., Ltd.	Haifa, Palestine	Crude Plant	124,000 bbls.	.....	.....	By 1950	Lummus, Kellogg	Lummus	Badger
Creole Pet. Corp.	Amuay Bay, Venezuela	Cat Cracking	60,000 bbls.	$70 million	Under Constr.	1950	.....	.....	.....
Creole Pet Corp.[3]	Molata Field, Venezuela	Stabilization and Repressuring Plants	.....	.....	.....	.....	.....	.....	.....
Creole Pet. Corp.[3]	El Roble Field, Venezuela	Repressuring	.....	.....	.....	.....	.....	.....	.....
Egyptian Government	Egypt	Crude Plant	.....	.....	.....	.....	.....	Lummus	Lummus
Imperial Oil, Ltd.	Montreal, East, Canada	Cat Cracker, Cat Poly	33,000 bbls.	.....	Under Constr.	Summer '48	.....	S.O.D., Kellogg	Kellogg
*Imperial Oil Ltd.	Edmonton, Canada	Thermal Crack, Crude Dist.	6,000 bbls.	.....	Under Constr.	Fall, 1948	.....	Barnes	Barnes
Imperial Oil, Ltd.	Ledue field	Gas absorption Plant	.....	$1.5 million	Projected	.....	.....	.....	.....
Industrie Chimiche Italiane Petroli	Mantau, Italy	Crude	4,500 bbls.	.....	.....	.....	.....	.....	.....
Mene Grande	West Guara, Venezuela	Repressuring	.....	.....	.....	.....	.....	.....	.....
Petro Carbon, Ltd.	Partington, Eng.	Cat Cracker, Lube & Reform	20,000 bbls.	.....	.....	.....	.....	.....	.....
Petro Chem., Ltd.	Partington, Eng.	Petro Chemicals	.....	.....	.....	.....	.....	.....	.....
Petroleos Mexicanos	Cindad Madero	Crude unit	25,000 bbls.	$2 million	Under Constr.	July, 1948	None	Own Staff	Own Staff
Petroleos Mexicanos	Salamanca, Mex.	Crude, gasoil Cracking	30,000 bbls.	$12 million	Contract let	1949	U.O.P.	McKee	McKee
Porto Marghera Rfg.	Venice, Italy	Enlarging	.....	.....	.....	.....	.....	.....	.....
Shell Chemical Corp.	Thornton, Eng.	Petro-chem	.....	Several millions	Under Constr.	.....	.....	.....	.....
Shell Chemical Corp.	Shell Haven, England	Petro-chem	.....	.....	.....	.....	.....	.....	.....
Shell Chemical Corp.	Stanlow, Eng.	Petro-chem	.....	.....	.....	.....	.....	.....	.....
Shell Oil Co (Venesuela)	Cardon, Venesuela	Complete Refg.	.....	.....	Under Constr.	.....	.....	Lummus	Lummus
Shell Oil Co.	Heysham, Eng.	Remodel & redesign Crude Plant	30,000 bbls.	.....	Under Constr.	Early '48	.....	.....	.....
Standard Oil Co. of B.C., Ltd.	Vancouver, B.C.	Asphalt Plant	600 tons	$500,000	Under Constr.	May, 1948	.....	Ralph M. Parsons	Own Staff.
Stora Kopparberg	Sweden	Crude, Ther. Cracking	.....	.....	.....	.....	.....	Lummus	Lummus

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REFINERY AND OTHER PLANT CONSTRUCTION (Continued)

COMPANY	Plant Site	Project	Daily Capacity	Estimated Cost	Status	Probable Completion	Licensor	Engineering	Contractor
Trent Oil Prod. Development Co.	North Shields, England	Crude Plant	11,500 bbls.	$4 to $8 million	Plans approved	.....	.....	.....	.....
Yacimientos Petroliferos Fiscales Bolivianos	Sucre, Bolivia	Crude Plant	3,000 bbls.	.....	.....	.....	.....	Foster-Wheeler	Foster-Wheeler
Yacimientos Petroliferos Fiscales, Bolivianos	Cochabamba, Bolivia	Topping & Reforming Units	5,000 bbls.	.....	.....	1949	.....	Foster-Wheeler	Foster-Wheeler
YPF Brasil.....	Santos, Brasil	Complete Refg.	20,000 bbls.	$30 million	Projected	.....	.....	.....	.....

NOTE—Atlantic—The Atlantic Refining Co. Badger—E. B. Badger and Sons. Barnes—W. M. Barnes & Co. Bellows—W. S Bellows Co. Braun—C. F. Braun Co. Cat. Constr.—Catalytic Construction Co. Clark's—Clark's Construction and Engineering Company. Dresser—Dresser Engineering Corporation. Fish Eng.—Fish Engineering Corp. Fluor—Fluor Corporation. Gaso. Plants—Gasoline Construction Co. Houdry—Houdry Corp. Hudson—Hudson Engineering Corp. Hydro Res.—Hydrocarbon Research, Inc. Jefferson—Jefferson Chemical Co. J & L Constr.—Jones & Laughlin Construction Co. Koch Eng—Koch Engineering Co. Kellogg—M. W. Kellogg Co. Leon Constr.—Leonard Construction Co. Lummus—The Lummus Co. Mason—Mason Brothers Construction Company. McKee—Arthur G. McKee Co. Old Dutch—Old Dutch Refining Company. Parsons—Ralph M. Parsons Co. Phillips—Phillips Petroleum Co. Pet. Eng.—Petroleum Engineers, Inc. Pritchard—J. F. Pritchard & Co. Process Eng.—Process Engineers, Inc. Ref. Eng. Co.—Refinery Engineering Co. S. O. D.—Standard Oil Development Co. Stearns-Roger—Stearns Roger Manufacturing Co. Texaco—Texaco Development Co. Warren—Warren Petroleum Corporation. Westheimer—Westheimer Rigging and Heavy Hauling Company. Wohfeld—Wohfeld Construction Co. [1] Jointly owned by Cities Service Oil Co. and Continental Oil Co. [2] Jointly owned with Pantepec Oil Co. [3] Jointly owned with Pantepec Oil Co. [4] Joint project with Cities Service Oil Co., The Carter Oil Co., Sinclair-Prairie Oil Co., with Warren operating plant. [5] Kerr-McGee Oil Industries, Inc., interest.
* First appearance in tabulation.

UOP Gets Contract for New Unit for Rock Island

Rock Island Refining Corporation, Indianapolis, has entered into an agreement with Universal Oil Products Company, Chicago, for the design and engineering of a fluid catalytic cracking unit to be installed in the Rock Island refinery at Rock Island (near Indianapolis) Ind.

This unit is to be designed for a capacity of 5000 barrels per day of Mid-Continent type gas oil at 54 percent conversion and is essentially the same type unit which Universal designed, and which has been on stream at the refinery of Aurora Gasoline Company, Detroit, since the summer of 1947.

The Rock Island unit will be so designed that it will utilize to the fullest extent the already existing thermal cracking unit and gas recovery system, it was explained. Feed preparation equipment for supplying charge to the cat cracker is already in place and will be a part of the new fluid unit.

New Shell Chemical Plant For Petroleum Additives

Construction is under way on a new Shell Chemical Corporation plant at Martinez, Calif., to manufacture "petroleum addititives" chemicals which are added to lubricating oils and other petroleum products to increase their life and performance.

Ralph M. Parsons Co. is the contractor for engineering and construction of the additives plant. It is scheduled for completion late this year.

United to Build Second Carthage Gasoline Plant

A second gasoline plant at Carthage, Texas, will be built by United Gas Corporation, it was announced last month, this being a part of United's current $14 million expansion program.

The plant will have a capacity of 100,000 mcf daily and is expected to be ready for operation in February, 1949. Hudson Engineering Corporation will handle the job, the cost of which will be about $3.5 million.

Texaco Expansion Work Progresses at Pt. Arthur

The Texas Company's expansion of lubricating oil and wax manufacturing facilities at the Port Arthur, Texas, refinery will include atmospheric crude topping and vacuum pipe still, solvent dewaxing unit, furfural refining unit and auxiliaries. It is expected that the work will be completed during the latter half of 1948. This completion will increase present plant capacities about 2500 barrels per day finished lubricating oils.

Foster Wheeler Corporation is handling the construction.

Midland Cooperative Enlarging Its Refinery

The Refinery Engineering Company of Tulsa and Chicago will supervise the new construction scheduled to increase capacity of the refinery and cracking units of Midland Cooperative Wholesale Company's Cushing, Okla., refinery, at a cost of $500,000, according to The Refinery Engineering Company, which will supervise construction. Walter Switzer has been named resident construction head of the project.

When completed, the plant will have a capacity stepped up from its present 4200 to 6500 barrels daily. Cracking plant facilities will have been enlarged 25 percent to make possible the processing of 2000 barrels of charging stock per day.

Sun Plans Gasoline Plant In Starr County, Texas

Clearing of a site in Starr County, Texas, for a natural gasoline plant equipped to process 25,000 mcf gas daily was announced last month by Sun Oil Company. Bids have been asked for actual construction.

The plant, expected to be in operation by April 1, 1949, is designed to recover 80 percent of the propane and virtually all the butane and natural gasoline from the casinghead gas produced in the Sun, North Sun, North Rincon and Garcia fields.

Super-Tankers Ordered

Two more super tankers with a capacity of 228,000 barrels each were ordered last month by Standard Oil Company (New Jersey), increasing to six the number of such tankers contracted for recently.

Each of the six vessels will have a capacity about 70 percent greater than the T-2 type tankers constructed during the war. The total capacity of all the ships will aggregate 57,456,000 gallons.

Garden City Selected as Site for Synthesis Plant

Garden City, Kan., has been selected by Stanolind Oil and Gas Company as the site for its synthesis plant in the Hugoton gas field. Soil surveys and water exploration tests will start immediately as will site preparation. Major construction will begin in the fall with completion set for January, 1950.

Garden City community leaders have assured the company that adequate rental housing for the permanent personnel of the plant will be provided and that the community will acquire the unused Army air base 12 miles from town for housing construction workers.

Stanolind's plant will take natural gas from the Hugoton field, some of which is not desirable for sale because of its low Btu. content, and will convert it into synthetic gasoline, diesel fuel, fuel oils, and useful chemicals.

The gasoline and fuel oils produced will be marketed largely in the Kansas area. The chemicals produced will be distributed by U. S. Industrial Chemicals, Inc.

As part of the project, the largest oxygen plant in America will be built at the site to supply oxygen for the process.

This Garden City project will include a natural gasoline plant to extract liquefiable hydrocarbons such as gasoline, butane, and propane from the natural gas; the synthesis plant itself including the oxygen production unit; a chemical refining unit, etc.

Sinclair Completes Part of Marcus Hook Expansion

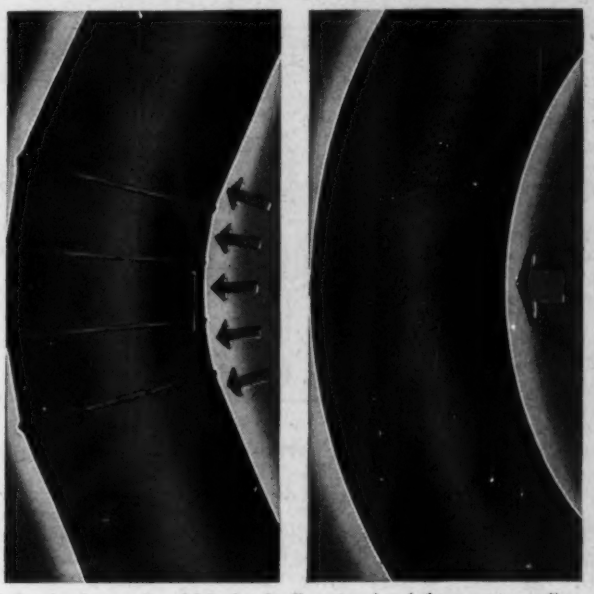
Sinclair Refining Company last month completed installation of a new catalytic cracking still at Marcus Hook, Pa., a unit capable of converting 25,000 barrels daily of charging stock into gasoline, heating oil, thermal cracking stock and gas. The installation is designed to permit flexibility in the yields of the several products, and will have a maximum daily gasoline producing capacity of 15,000 barrels and a maximum daily heating oil capacity of 6000 barrels.

Gas produced will go through a polymerization plant, now under construction, in which portions will be converted into gasoline.

Also added to the Marcus Hook plant are a new boiler house, a boiler feed water treating plant, an enlarged water supply plant and two vacuum stills.

NEW KELLOGG PROCESS MAKES GIANT PIPE BENDS *In one piece!*

Avoids corrosion and erosion due to stress intensification . . . applicable to pipe up to 8 feet in diameter



In pipe larger than 34 inches in diameter, bends have customarily been made by welding several segments together (left). The non-uniform contour of such welded bends increased erosive effects of flowing material because of the abrupt changes in direction of flow. In addition, a serious phenomenon, known as stress corrosion developed at or near weld seams. The smooth contour of the new Kellogg one-piece bends reduces the factors of erosion and corrosion to a minimum.

M. W. KELLOGG engineers have perfected a method and apparatus for bending, *in one piece,* pipe large enough to hold a limousine!

By eliminating the need for welded segments, the new process permits smoother, more uniform flow of fluids . . . reduces erosion and wear to a minimum, particularly where powdered catalysts are utilized.

Through an ingenious application of heat and hydraulic pressure, large diameter pipe is bent in continuous incremental stages without buckling or creasing. The process is adaptable alike to small and large diameters, to thin and thick-walled pipe, and to short and long radius bends.

Exhaustive tests and analyses of 50-inch pipe, currently being fabricated in the Kellogg Shops, prove that the new weldless, one-piece bending process produces results superior to any previously obtained.

It's this constant striving for improvement that accounts for Kellogg's position in petroleum engineering . . . the reason why so many of the world's foremost petroleum companies turn to Kellogg when top performance is all-important.

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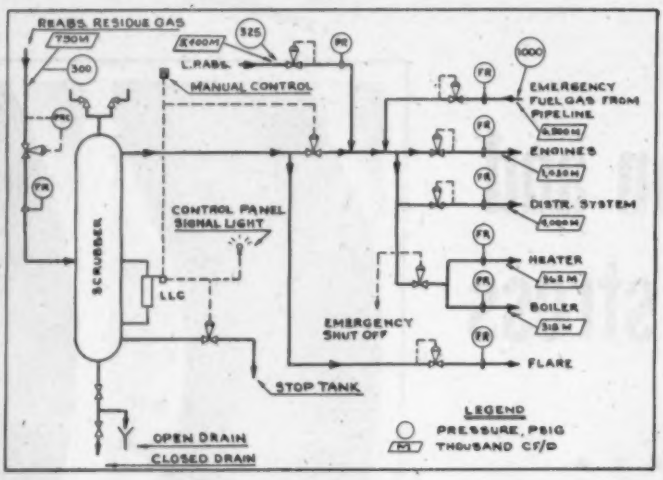
OPERATION HOW TO DO IT MAINTENANCE

DETAILS OF—

Plant Fuel Gas Scrubber Control System

A FUEL gas scrubber and distribution system to supply an uninterrupted flow of gas with a minimum of maintenance is of prime importance to efficient gasoline plant operation. The operation of these systems is aided by proper application of controlling instruments. In the diagrammed system below, which is installed in the Vanderbilt gasoline plant of Magnolia Petroleum Company, the instrumentation method used requires that the system receive routine inspection only, and eliminates the need for manual control except during an emergency.



The plant has 750 Mcf daily of residue gas from a reabsorber which goes into the fuel system after passing through a fuel gas scrubber to remove entrained liquid. A liquid level controller prevents carry-over of liquid from the scrubber into the fuel system by closing a valve in the gas line and opening a valve in the line to a "stop" tank to drain off the liquid. The controller also closes a switch for a signal light on the control panel. In such an emergency gas from the scrubber is flared until the liquid has been drained from the scrubber. If the drain valve should not operate properly in draining the liquid, the operator has a manual control that closes the gas line valve and prevents liquid carry-over until the liquid is drained.

Supply of Emergency Gas

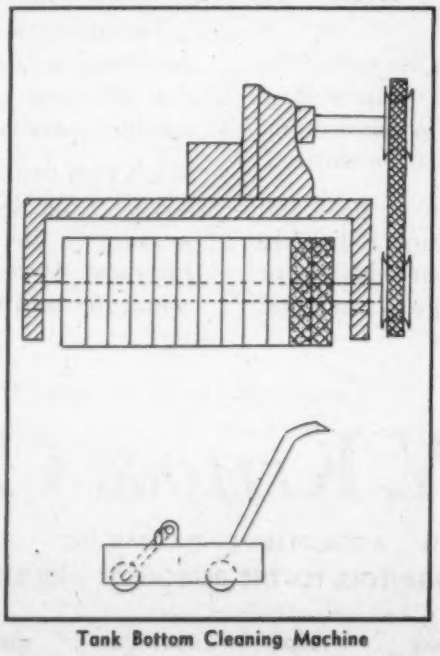
As the total fuel requirements for the plant amounts to about 6110 Mcf daily, the remainder of the fuel gas is supplied by gas from the low pressure absorber residue scrubber, 5400 Mcf daily. This gas enters the system through a pressure regulator on the downstream side of the level controlled gas line valve.

When the fuel gas scrubber is shutdown, the volume of gas available from the low pressure absorber residue scrubber is not sufficient for fuel requirements. This difference is made up with gas from the gas transmission pipe line, operated at 1000 psig. with 6500 Mcf daily available for use as fuel. The line provides a floating supply of emergency fuel gas.

A pressure regulator maintains the pressure in the fuel system with make-up gas from the transmission line. The excess gas in the system is flared through the flare line pressure regulator, which has an opening pressure slightly above the pressure maintained in the system. Each individual gas line is equipped with a pressure regulator to maintain the required pressure. Flow meters on each inlet and outlet line of the system record the volumes of gas.

Machine Devised to Clean Tank Bottoms

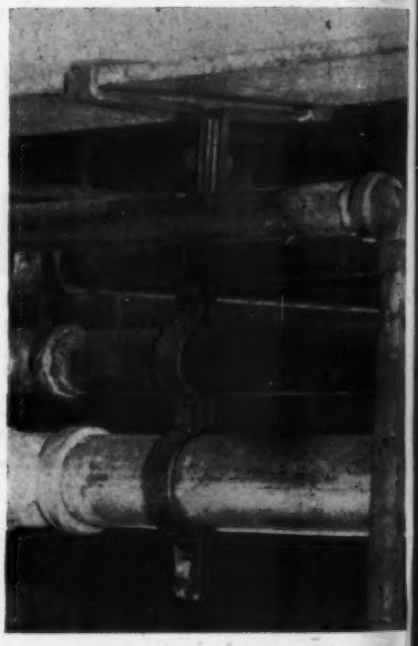
UTILIZING a small air-driven motor, pulleys, belts, and circular brass wire brushes, Robert Hines, tank foreman with Humble Pipe Line Company at Longview, Texas, fabricated a tank bottom cleaning machine which facilitates in the cleaning of the bottoms of oil storage tanks.



Tank Bottom Cleaning Machine

The device is similar in operation and appearance to a floor sanding machine. The motor is mounted slightly forward on the chassis and is connected to the brushes by V-belt and pulleys. The weight of the motor provides sufficient pressure upon the brushes to assure satisfactory cleaning action. The 12 brass wire brushes mounted on the axle clean a path approximately a foot wide. In using the air-driven motor and brass brushes, the possibility of sparking is precluded.

The motor can be removed for use elsewhere when the tank cleaning operation is completed.



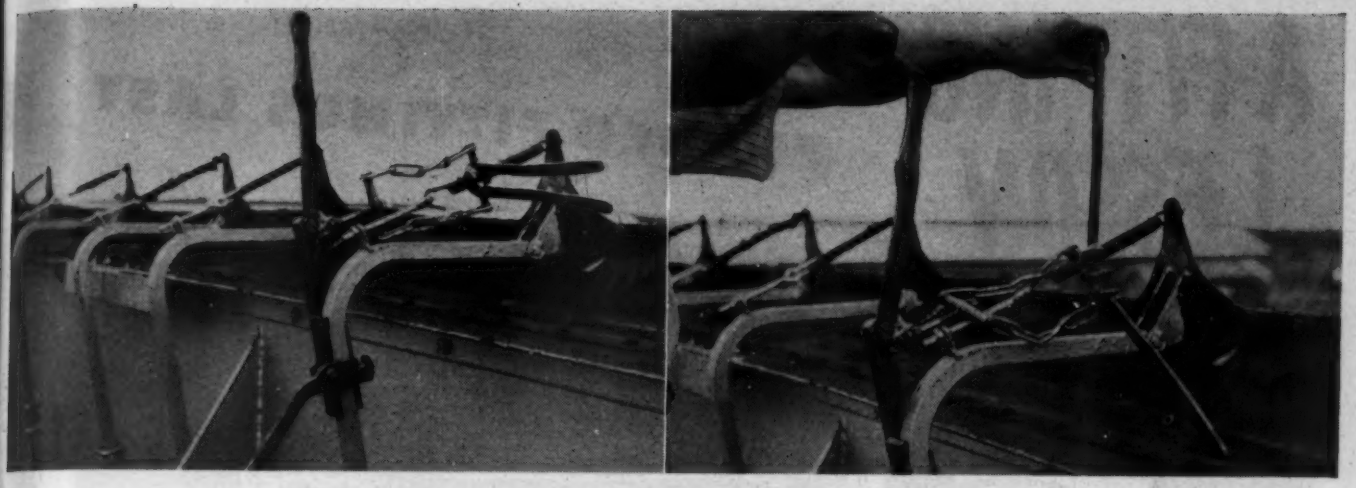
HOW—

Pipe Can Be Hung From Overhead I-Beams

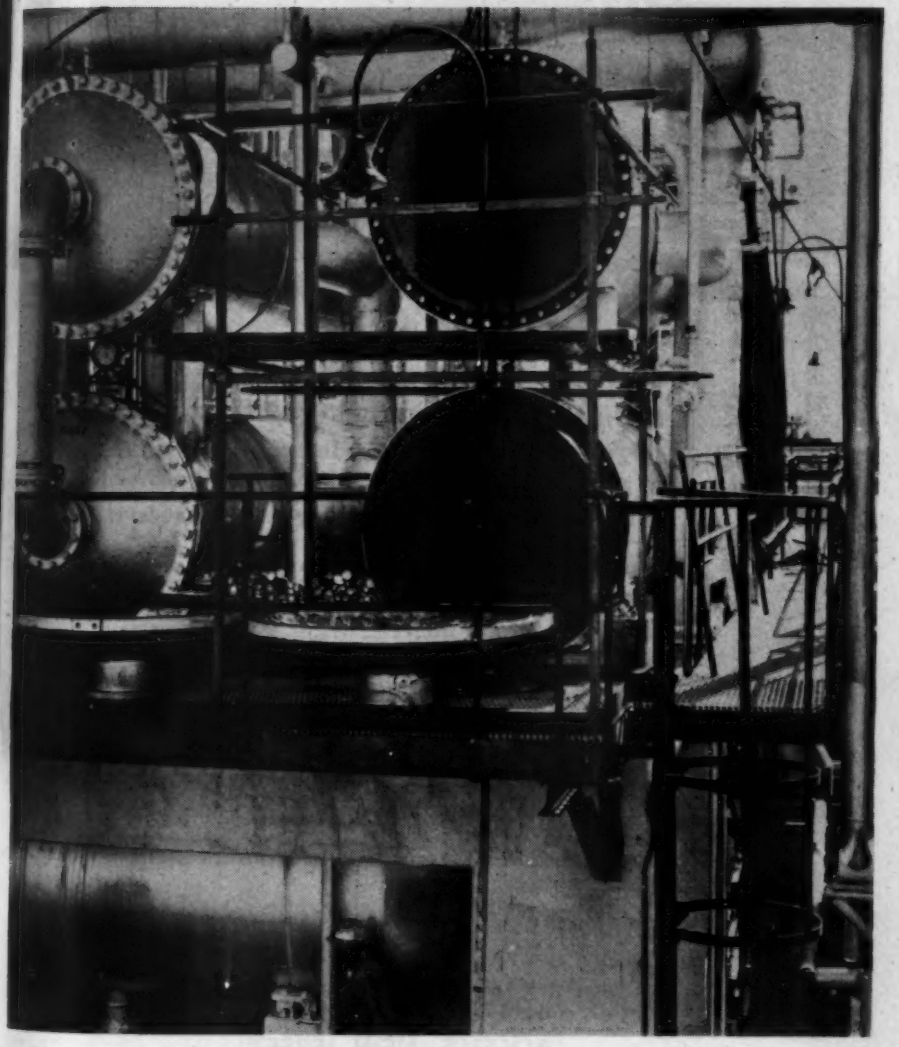
A HANGER that clamps to the overhead I-beams of structural framed buildings offers a convenient method for supporting overhead lines. Several lines can be supported by each hanger and if hangers are not placed too far apart, they will prove effective as vibration dampeners.

The hanger consists of four parts, two clamping jaws and two straps which fit snugly around the various sizes of piping installed. Two bolts hold the assembled hanger together.

The installation of pipe in this manner presents a neat and orderly appearance. Simplicity of construction adds to the effectiveness of the hanger.



Pictures above show how device works for removing springs from floating roof tanks.



HOW TO—

Catch Wash Effluent from Condensers

WHEN cleaning condensers located above grade, the maintenance department of the butadiene plant of Cities Service Refining Corporation at Lake Charles, La., utilizes a splash-shield pan similar to the condenser head to catch the wash effluent. The dirty effluent is carried to the plant sewer system via a fire hose attached to a drain coupling at the bottom of the pan.

The splash-shield pan is made of light weight steel and bolts to the open end of the condenser. **(See photo above.)** A gasket prevents the collected effluent from leaking.

HOW TO—

Remove Springs from Floating Roof Tank

(See photos above)

THERE are 160 upper coil springs on a typical 55,000 barrel floating roof tank. The springs must be removed periodically for greasing to prevent rust and corrosion. To remove each spring an eye-bolt nut is loosened approximately 3 inches to allow the spring hook to be slipped off. Often the threads are rusted and the eye bolts are broken in trying to remove the nuts.

A spring jack for removing these coil springs, devised by Mike Dobray of Humble Pipe Line Company, is pictured herewith.

Dobray's jack slips on the hanger bar and connects to the spring so that when a lever is pulled down the spring is stretched and can be lifted from the eye bolt without the eye-bolt nut being loosened.

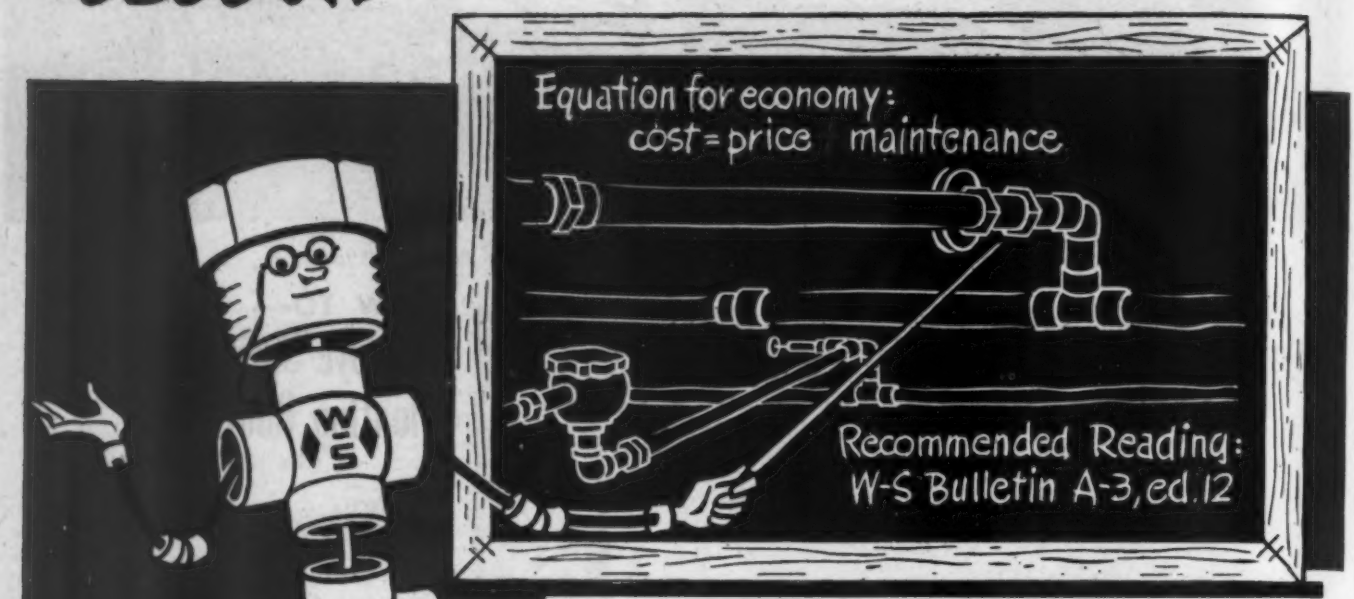
Modified Stripper Is Big Improvement

ONE-THIRD of the gas-oil feed stock for the three Fluid catalytic cracking units at Cities Service Oil Company's Lake Charles, La., refinery is prepared by propane de-asphalting reduced crude. The de-asphalting operation is controlled to produce a very heavy de-asphalted gas-oil in order to maximize catalytic cracking unit feed stock availability. Only 2 percent of the crude is normally yielded as asphalt. As a result of such operation the de-asphalted residual gas-oil tends to produce high catalytic coke yield and contaminates cracking catalyst rapidly.

Experience at Lake Charles has proven that with conventional fluid unit operating procedures about 10,000 barrels per day of this heavy residual gas-oil can be blended into the total feed stock charged to one cat cracker. As the refinery crude rates were stepped up the production of this gas-oil increased to 20,000 barrels per day which made it necessary to blend the de-asphalted gas-oil into the feed stock of two catalytic units,

A FITTING LESSON

FORGED FITTINGS LAST



Right Professor. ***THE MAJOR COST OF A FITTING IS THE COST OF ITS REPLACEMENT.*** And the life of one *forged* steel fitting is many times greater than that of any other fitting. That's why W-S *forged* steel fittings cost less in the long run, for any industrial service.

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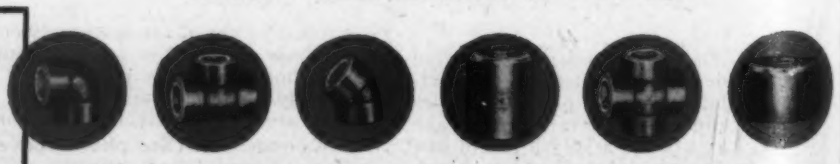


SOLD THROUGH LEADING DISTRIBUTORS

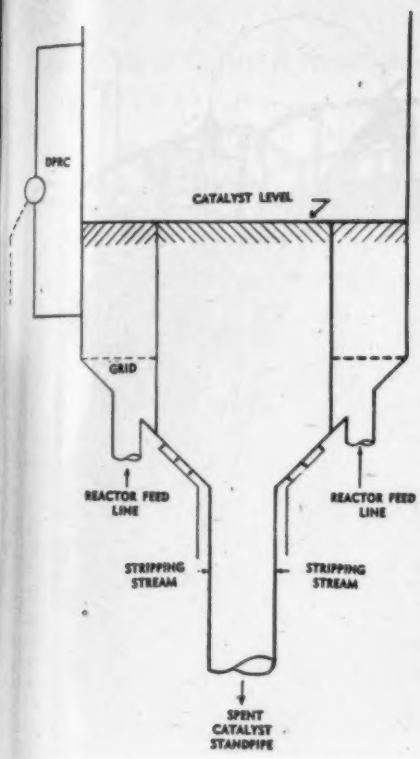


FIGURE 1
Single-Level Operation

thus increasing the rate of contamination on two beds of catalyst.

Fortunately, however, the Lake Charles operators developed a modified stripper which resulted in their again being able to charge the de-asphalted gas-oil production to one unit. Figure 1 shows the conventional down flow unit stripping arrangement and Figure 2 indicates the Cities Service modification. The magnitude of the improvement is shown by the following data:

Comparison of Conventional and Two Level Stripper Operation

Feed Stock—BPSD	Conventional	Two Level
De-asphalted Oil	10,300	20,205
Distillate Gasoil	20,923	10,087
Total Feed	31,223	30,292
Carbon Residue on Feed Stock, Weight percent	0.95	2.06

Several other major companies have met with similar success with the two level operation. It is evident that the cost of making the conversion is negligible.

"Coin Your Idea"

Contributions (with photos or sketches) to this "How To Do It" Department are invited.

Address the Editor

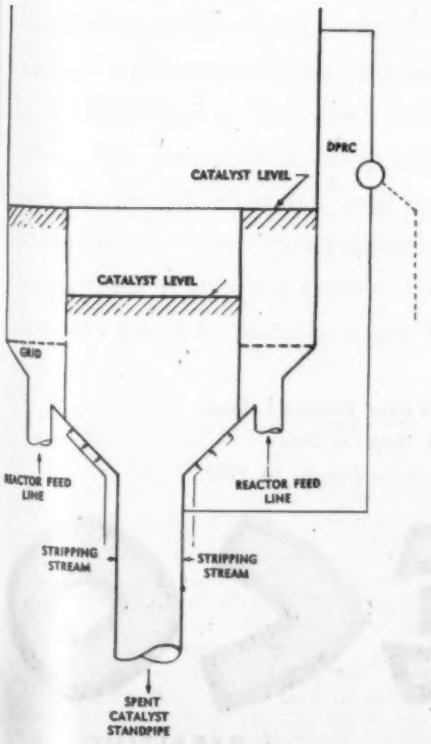


FIGURE 2
Two-Level Operation

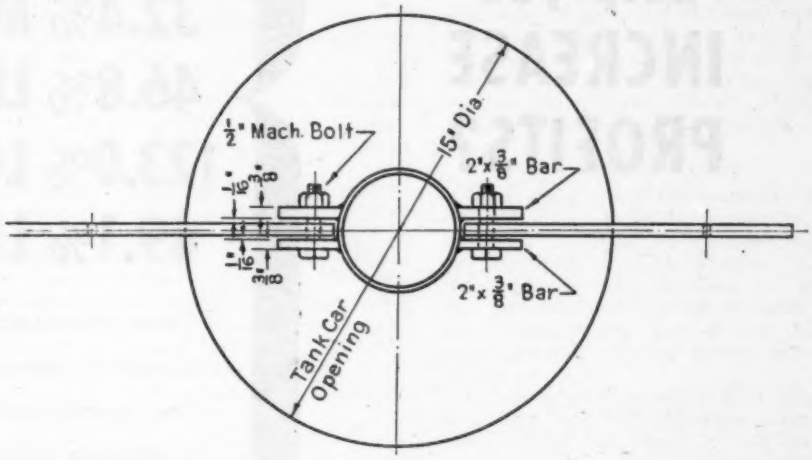
HOW TO—

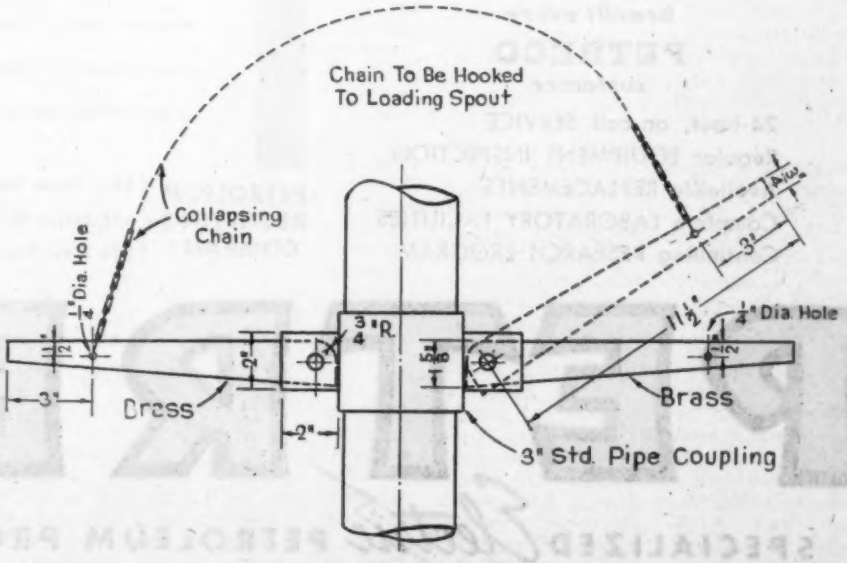
Solve One Problem of Tank Car Loading

IT IS more or less normal practice to load tank cars carrying non-volatile petroleum liquids by means of a flexible hose which is inserted into an open manhole in the car dome. This practice has sometimes resulted in accidents where a loading line would come out of the tank car dome and spill petroleum products over the car and over operating personnel.

One satisfactory solution has been worked out by Ralph Williams of Wood River Oil and Refining Company, Inc., located at Hartford, Ill. The device, shown in the accompanying sketch, consists of a pair of folding arms attached to a clamp which in turn can be bolted to the loading line. The arms are so designed that, when the line is inserted in the tank dome, they lock in an extended position and prevent removal of the line. A short piece of chain looped between the two projecting arms serves to pull these arms back into a position close to the loading line when it is desired to remove the line.

(See sketches below.)





Design of Tank Feed Line Hook

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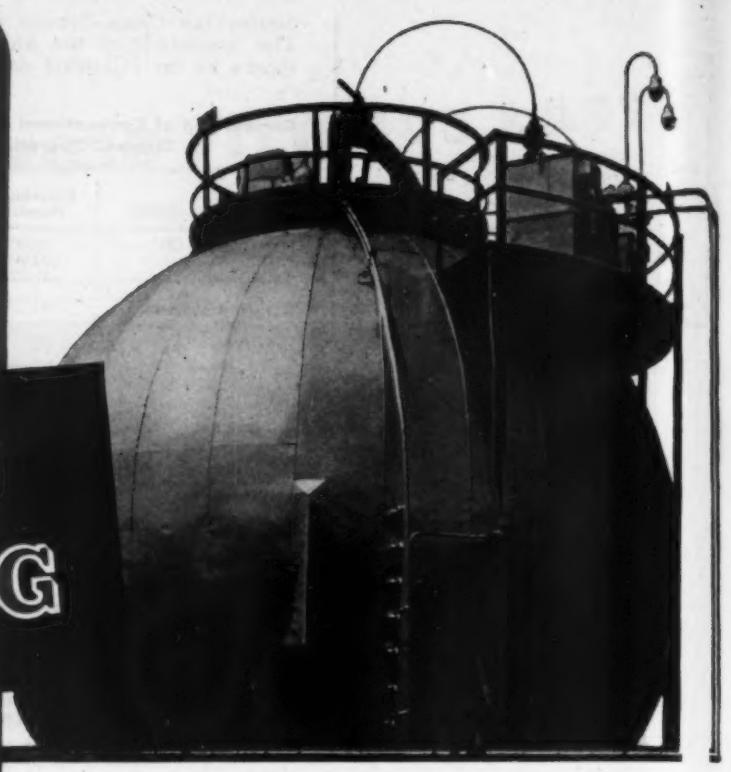

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Born in a Critical Period—

Natural Gasoline Group Boasts Record of Accomplishment

FORMATION of what is now the Natural Gasoline Association of America resulted from a critical period for the casinghead gasoline maker back in 1921. Prices were so low that something had to be done. One of the first things decided upon was a set of specifications. The necessity for this later was described by H. T. Ashton, now manager of the Lubrite Division, Socony-Vacuum Oil Company, in his paper before the annual meeting in 1926:

"Even in 1921 a check of the Mid-Continent area revealed that there were more than 100 different grades of raw casinghead gasoline and blends being manufactured. This was an impossible situation and the association went to work to unsnarl this tangled mass of products."

Earlier M. W. Welch, first president of the association, had outlined the plight of the makers of the lighter hydrocarbons with:

"We believe that the greatest good would be accomplished by commencing at and with the internal and intrinsic problems of our industry. We felt that if we set our house in order the satisfactory solution of the external or marketing problems would follow and that our commodity would logically and naturally take its true position in the price scale of petroleum problems. Instead of playing with the price situation, which was threatening to consume us, we tried to make our structure fire proof."

Specifications Approved

The organization was formed April 21, 1921 when a few manufacturers met in Tulsa and its name was the Association of Casinghead Gasoline Manufacturers. By July of that year the association had a set of specifications approved and the name had been changed to the Association of Natural Gasoline Manufacturers.

In 1926 the name was changed again to the Natural Gasoline Association of America.

Nineteen twenty-six also marked the beginning of the 10-year research program through which natural gasoline gained its rightful place in the family of petroleum products. Dr. G. G. Brown of the University of Michigan appeared on the program at the fifth annual meeting in Tulsa and his connection with the association began shortly afterward.

His first annual report a year later brought the statement that as much as 50 percent natural gasoline blend could be used in motor fuel to advantage. Further he observed that the first move to that end was that he "had to sell the natural gasoline manufacturers themselves."

That the natural gasoline manufacturers were willing students is evident from the growth of the association. Currently some of its committee meetings have attendance to exceed the total registration of the casinghead gasoline manufacturers who met in desperation in Tulsa in 1921. Activity of the association has broadened into investigations of the uses of its products, an instance of which is the study of causes and correction of vapor lock, known officially as the Pawhuska road tests, which also provide data on many other phases of fuel performance in automobile operation. More recently the association has undertaken the study of causes of corrosion of tubing and casing in distillate wells.

Year after year the association has kept the industry informed on advancements and refinements in plant operating methods. The economic situation has been analyzed as conditions have changed in relation to the lighter hydrocarbons. The more recent turn of the industry to petrochemical processing has resulted in programs to include these developments. Likewise the liquefied petroleum gas expansion has demanded a place on association programs.

Since formation of the California Natural Gasoline Association in 1925 the two organizations have worked together in many efforts.

Although the annual meeting of the Natural Gasoline Association of America is its high point in activity, the work of the association goes on constantly trough committees. Here specifications are revised; plant operating problems studied; cooperation with other associations furthered and education of its own members helped. These committees assist the affiliated Natural Gasoline Supply Men's Association in editing and producing the Technical Bulletin, which has come to be a standard reference work for the whole petroleum industry.

No better evidence of the position the association holds within the petroleum industry could be offered than the registration report from the twenty-seventh annual meeting in Fort Worth last month when 952 persons registered.

Meeting Development

In the early years of the association, one-day sessions were held in various oil cities and were called "Plant Operator's Meetings." The programs were arranged for these men and filled the need for the exchange of information on the wide variety of operating techniques, safety in handling volatile gases and even what to do with the products manufactured, which was needed by practically everyone in the business. As the industry became more stabilized and fundamentals of good plant practice were better understood, these meetings were discontinued. Later it was realized that men in the plant and field have advanced as fast as the industry, but their problems also multiplied as the industry advanced. Annual conventions were of necessity too crowded with discussions of industry policy and broad economics to permit inclusion of many operating papers. Also, the large territory over which the industry is spread, prevented many plant and field men from attending the annual convention.

The Panhandle Plains regional meeting, first of the regional meetings, was held in Amarillo, Texas, last December and drew an attendance of 300. The second was the Gulf Coast regional meeting in Corpus Christi, Texas, in January with nearly 300 present.

Hanlon Award

Each year the NGAA selects an individual who has made outstanding contribution to the natural gasoline industry as recipient of The Hanlon Award. Twelve men have received the award, the highest honor in the natural gasoline industry and one of the ranking merit awards of the entire petroleum industry.

Donor of the award is E. I. Hanlon, president of The Hanlon Company, chairman of the board of the National Bank of Tulsa, and a pioneer of the natural gasoline industry.

Officers for 1947-1948 include President, C. R. Williams, The Chicago Corporation and four vice presidents: Henry M. Brown, Midland Gasoline Company; R. D. Gibbs, Union Oil Company of California; James E. Pew, Sun Oil Company and Albert H. Weil, United Gas Pipe Line Company.

The board of directors comprises: John B. Atkins, The Parade Company, Shreveport, La.; Phil D. Baker, The Carter Oil Company, Tulsa; C. E. Brock, Kansas Power & Light Company, Salina, Kan.; George P. Bunn, Phillips Petroleum Company, Bartlesville, Okla.; Henry M. Brown, Midland Gasoline Company, Houston; J. H. Dunn, The Shamrock Oil and Gas Corporation, Amarillo, Texas; R. D. Gibbs, Union Oil Company of California, Los Angeles; Allen T. Givens, Coltexo Corporation, Monroe, La.; T. R. Goebel, Shell Oil Co., Inc., Houston; F. B. Haverfield, Continental Oil Company, Ponca City, Okla.; J. A. LaFortune, Warren Petroleum Corporation, Tulsa; M. R. Lents, J. S. Abercrombie Com-

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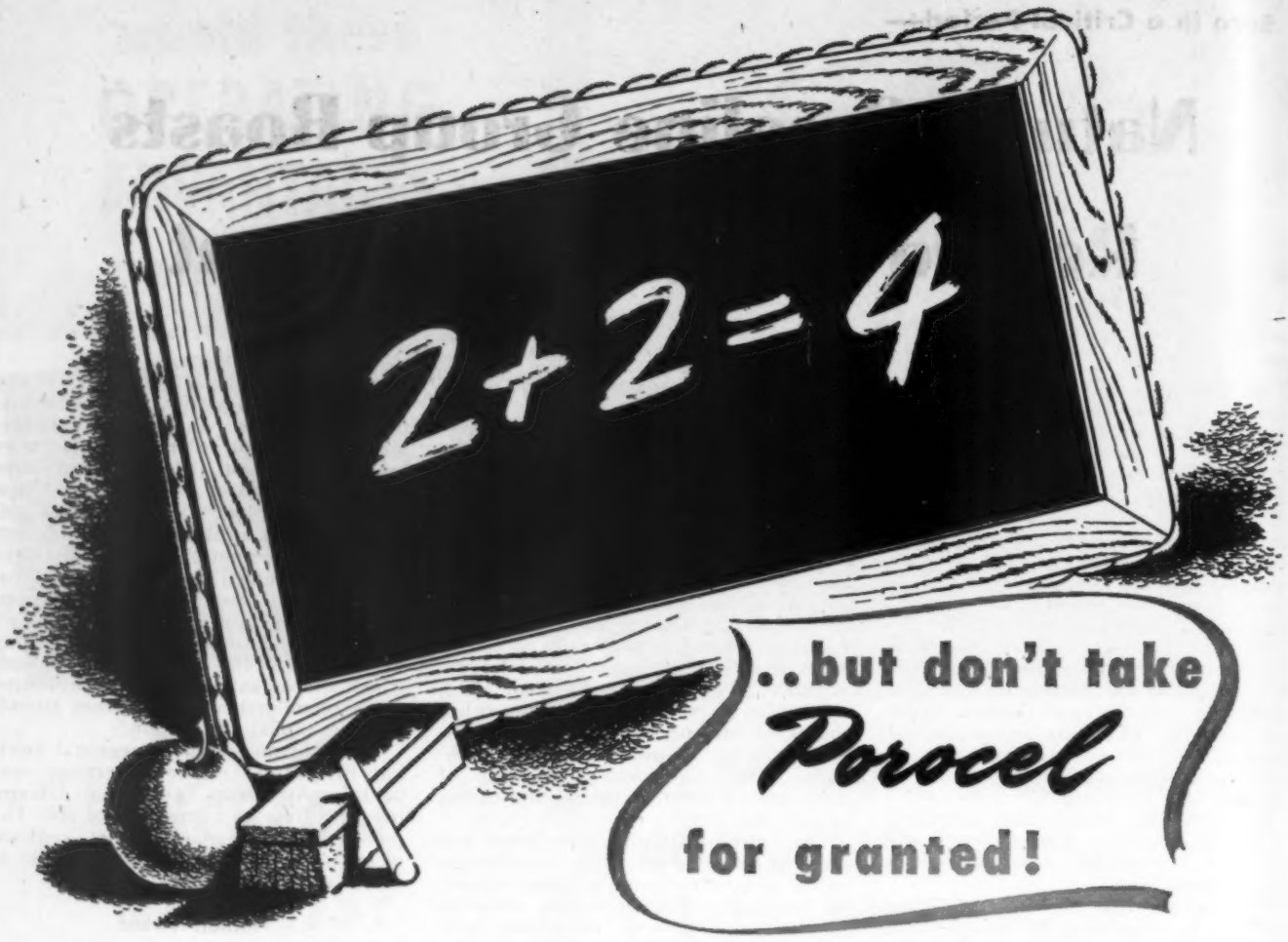

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California Has Own Association

FORMED in 1925 for the purpose of standardizing technical procedures for the natural gasoline industry in the West, the California Natural Gasoline Association now has some 600 active members and received the support of some 200 oil, natural gasoline, gas equipment and engineering firms. Its industry-wide recognition as a technical nonprofit educational organization is the natural result of 23 years of effort to bring about high standards of test methods and gas measurement procedure, and to serve as a medium for the dissemination and exchange of information. Monthly meetings are held in Taft, Ventura and Los Angeles.

The aims of the association are not the work of any one man; rather the procedure and rules under which CNGA works grew out of its formative years. Principles were laid down in the first few years which have been maintained to insure continuing service to the industry. These basic rules may be summarized:

(1) The CNGA discharged a responsibility to the personnel of the natural gasoline industry in the matter of their *education* to better fit them for their work, acquaint them with some of the underlying technical facts necessary to the efficient execution of their daily duties, and to provide meetings where all engaged in making natural gasoline can meet and discuss mutual problems.

(2) *Standardization.* CNGA has arrived at standards of procedures for metering and testing for natural gas and gasoline, which are acceptable throughout the industry by buyers and sellers. The legal advantage of possessing a recognized procedure for testing can best be appreciated by those who have attempted to detail a particular test method in their contracts. It obviously is simpler to merely include a paragraph stating the test procedure shall be as detailed in the appropriate CNGA Bulletin.

(3) The organization supplies the oil industry with *reliable information,* showing the statistical position of the natural gas and natural gasoline business from month to month. Companies furnished data on a strictly confidential basis, the result being published by fields only.

(4) A guiding NGAA principle has been that the group, primarily technical in their interests, should approach their problems from the viewpoint of engineers, and should maintain strict professional *ethics* in their contacts with commercial organizations.

These basic ideals have been helpful in preserving the association through difficult times and have been responsible largely for its excellent esprit de corps.

Problems Outlined

Committees are formed to carry on the never ending work of standardization. Technical members of these groups receive the cooperation of their respective companies in the matter of time and use of needed facilities. Problems requiring extensive research are not undertaken; rather the study of available methods of doing a job and the development and recommendation for adoption as a standard of the procedures thought best.

For the fiscal year of 1947-48 the technical committee has had certain problems assigned them which in part constitute their present objectives as follows:

Problem No. 1: Color Code Committee—As the result of discussion on the apparent necessity of a uniform code of marking pipe lines an examination is being made of codes and systems now in use.

Problem No. 2: Gas Measurement Committee— CNGA's present Bulletin TS-353 does not include the coefficients for orifice plates to be used in heavy and extra heavy pipe. These coefficients will be computed by the method given in TS-353 jointly with Pacific Coast Gas Association and the Southern California Meter Association as a supplement to TS-353.

Problem No. 3: Fractional Analysis Committee—This committee is considering three projects:

1) Possible revision of TS-411 relating to the dimensions of the condenser bath; 2) Make a comparison of the recent NGAA fractional analysis procedure with that outlined in CNGA Bulletin TS-411 to determine if any conflicts exist and if CNGA procedure comes within the specifications of the NGAA procedure; 3) Investigation and study of the silver bead corrosion test for LPG products.

In addition to these problems the CNGA technical committee actively cooperates with the ASTM and NGAA and API in the development of technical standards for industry.

During its 23 years, the CNGA has been actively and successfully engaged in bringing about a condition of technical uniformity within the industry. The publication of 15 tentative standards represents sound steps forward in the technical advance of the industry.

Officers and Directors

Officers for the fiscal year 1947-1948 include: President, F. J. Colton, Tide Water Associated Oil Company; Vice President, W. A. Kirk, California Southern Oil Company; and Secretary-Treasurer, E. R. Millett, Jr., California Natural Gasoline Association.

Directors follow: (asterisk indicates past presidents and members of executive committee, dagger indicates members of executive committee.) *M. L. Arnold, Richfield Oil Corporation; Coles B. Bason, Ingersoll-Rand Company; A. H. Bradford, West Coast Refining Company; †John G. Braun, C. F. Braun & Company; J. A. Campbell, J. A. Campbell Company; *L. V. Cassaday, Lomita Gasoline Company; F. J. Colton, Tide Water Associated Oil Company; D. A. Cushman, Fullerton Oil Company; *W. C. Dayhuff, California Research Corporation; F. F. Doyle, Pacific Gas & Electric Company; †W. Earl Dunn, The Fluor Corporation, Ltd.; *H. L. Eggleston, General Petroleum Corporation; †R. C. Enderly, Wilmington Gasoline Company; M. L. Fort, Pacific Lighting Corporation; R. E. Foss, Barnsdall Oil Company; *C. D. Gard, Union Oil Company of California; *R. D. Gibbs, Union Oil Company of California; *R. W. Heath, Signal Oil & Gas Company; J. C. Herron, Shell Oil Company, Inc.; †C. L. Hutchings, Tide Water Assoc. Oil Company; *M. W. Kibre, General Petroleum Corporation; W. A. Kirk, California Southern Oil Company; †W. H. Kraft, Honolulu Oil Corporation; *H. R. Linhoff, Richfield Oil Corporation; R. S. Lytle, operator; S. F. Magor, The Superior Oil Company; *P. S. Magruder, General Petroleum Corporation; †Geo. C. McLaren, Standard Oil Com-

Natural Gasoline Group

• CONTINUED FROM PAGE 163

pany, Houston; John F. Lynch, La Gloria Corporation, Corpus Christi, Texas; H. W. Manley, Barnsdall Oil Company, Tulsa; I. Earl Nutter, Hagy, Harrington & Marsh, Amarillo; R. N. Parks, Virginian Gasoline & Oil Company, Charleston, W. Va.; F. M. Perry, Cities Service Oil Company, Bartlesville, Okla.; James E. Pew, Sun Oil Company, Philadelphia; Paul M. Raigorodsky, Glen Rose Gasoline Company, Houston; T. L. Taggart, Standard Oil Company of California, San Francisco; F. H. Townsend, Lone Star Producing Company, Dallas; James W. Vaiden, Skelly Oil Company, Tulsa; W. H. Vaughan, Tide Water Associated Oil Company, Houston; Albert H. Weil, United Gas Pipe Line Company, Shreveport; C. R. Williams, The Chicago Corporation, Corpus Christi; and Evan A. Thomas, Cotton Valley Operators Committee, Cotton Valley, La.

William F. Lowe has been secretary of the association since 1934.

Supply Men's Group

SOMETHING unique in association activity is the functioning of the Natural Gasoline Supply Men's Association. Its one purpose is to provide entertainment at the annual meeting of the Natural Gasoline Association of America. Each year it has a brief business meeting at which officials are elected and reports made.

Then it stages the whole program for banquet and entertainment as fits the occasion. It has been providing this auxiliary service since 1928.

This association had its start when manufacturers of equipment sought means of promoting their own affairs while aiding the attendance of convention programs. The result was that manufacturers grouped their effort for one round of hospitality at each annual meeting.

Back in 1935 the supply association provided a Technical Bulletin, the popularity of which was so widespread that the fifth edition was necessary by 1946. Data for this bulletin are provided through cooperative effort of technical committees of the two associations and revisions are made as more data reach the status of acceptance in plant operation.

Officers of the association for 1947-48 are as follows: Roy R. Bush, Nordstrom Valve Company, president; J. N. McClure, Elliott Company, first vice president; L. L. Dresser, Dresser Engineering Company, second vice-president; Dallas S. Deems, American Air Filter Company, Inc., treasurer; and W. F. Lowe secretary (reelected).

pany of California; †Wm. Moeller, Jr., Southern California Gas Company; †H. E. Moulton, C. C. M. O. Company; *N. H. Mull, The Texas Company; H. W. Parmelee, Del Valle Gasoline Company; †H. M. Paulsen, The Norwalk Company; *W. W. Robinson, Jr., The Texas Company; J. G. Sargent, K. N. D. A.; W. D. Smith, Wilshire Oil Company; †George Sowards, The Ohio Oil Company; †J. B. Taylor, Signal Oil & Gas Company; †R. S. Tulin, Shell Oil Company, Inc.; *T. L. Taggart, Standard Oil Company of California; E. W. Walker, Western Gulf Oil Company; E. W. Webb, Continental Oil Company and D. M. Wolfe, Standard Oil Company of California.

Committee Chairmen

CNGA committee chairmen include:

Advisory Committee—W. A. Kirk, California Southern Oil Company.

By-Laws Revisions Committee—J. B. Taylor, Signal Oil & Gas Company.

Entertainment Committee—R. D. Raasch, The OFCCO Company.

Fall Meeting Committee— R. S. Ridgway, Standard Oil Company.

Finance and Budget Committee—R. W. Heath, Signal Oil & Gas Company.

Gas Report Committee—R. S. Tulin, Shell Oil Company, Inc.

Membership Committee—R. W. Todd, Pacific Lighting Corporation.

Program Committee—Paul Armstrong, Union Steam Pump Sales.

Publicity Committee—M. L. Arnold, Richfield Oil Corporation.

Technical Committees and sub-committees include:

Technical Committee—E. P. Valby, Richfield Oil Corporation; D. V. Johns, California Research Corporation, vice chairman.

Technical Advisory Committee—W. C. Dayhuff, California Research Corporation; R. R. Crippen, The Texas Company, vice chairman.

Gas Measurement Committee No. 1—Mark McIlvaine, C. C. M. O. Company.

Low Temperature Fractional Analysis No. 2—R. A. King, Union Oil Company.

Physical Constants Committee No. 5—George Horne, Richfield Oil Corporation.

Charcoal Test Committee No. 6—J. E. Adams, Signal Oil & Gas Company.

Liquefied Petroleum Gas Committee No. 9—J. R. Spencer, California Research Corporation.

Editorials & Revisions Committee No. 10—R. J. Allen, General Petroleum Corporation.

Orsat Analysis Method Committee No. 14—R. A. Montague, Standard Oil Company of California.

Piping Color Code Committee No. 15—R. F. Vincent, The Texas Company.

Westinghouse Research Now Headed by Hutcheson

Dr. John A. Hutcheson has been appointed director of the Westinghouse research laboratories, succeeding Dr. L. Warrington Chubb, who is retiring from active direction for reasons of health, but who will continue to serve in an advisory capacity. Dr. Hutcheson has been associate director for four years. During this period he directed the company's extensive wartime radar research program and later the formulation of plans for atomic energy development. He will remain active in both fields.

NGAA Convention

Big attendance at Fort Worth gathering hears interesting discussions

GORDON W. ALEXANDER, Associate Editor

THE 27th annual convention of the Natural Gasoline Association of America was held in Fort Worth, Texas, on March 24, 25, and 26 with between 950 to 1000 plant operators, design engineers, equipment men, and others concerned with technical developments within the natural gasoline and petrochemical manufacturing industry in attendance.

Highlight of the meeting was the presentation of the Hanlon Award to Colonel George A. Burrell, president, Atlantic States Gas Company, Inc., New York. Presentation was by NGAA president, C. R. (Dick) Williams, The Chicago Corporation, Corpus Christi, Texas.

Colonel Burrell is the twelfth recipient of the award, which is conferred each year for outstanding service to the natural gasoline and cycling industries. The citation read in part: "There is little doubt that this man was the first experimenter to make an exact analysis of natural gas. He conceived the type and character of equipment required, made and assembled it himself, and made it work. This was in 1915. It is scarcely open to question that this was one of the most important detail steps in the technological progress of the petroleum-gas field."

Officers Reelected

The officers now active in the association were reelected for the coming year as follow: C. R. Williams, The Chicago Corporation, Corpus Christi, Teaxs, president; Albert H. Weil, United Gas Pipeline Company, Shreveport, La., R. D. Gibbs, Union Oil Company, Los Angeles, James E. Pew, Sun Oil Company, Dallas, and Henry M. Brown, Midland Gasoline Company, Houston, vice presidents; and William F. Lowe, secretary-treasurer.

Featured papers presented at the sessions dealt with the increase in production of lighter hydrocarbons at lower costs, better utilization of these hydrocarbons, maximum possible recovery of natural gas, and conservation of the various hydrocarbon constituents of oil and gas reserves. These were timely and received with great interest due to the unprecedented demand for these products for use in liquefied petroleum gases markets and the manufacture of motor fuels.

The initial session of the convention was devoted to cycling operations with John F. Lynch, La Gloria Corporation, as chairman.

Plant Designer Paper

"Condensation of Hydrocarbon-Water Mixtures on Finned Tubes" by Donald L. Katz and Daniel C. Gillespie of the University of Michigan, presented data gathered from experiments to find the static retention of liquid on finned tubes as a function of the ratio of surface tension to liquid density.

Finned tubes provide extra surface on the outside, and prove valuable in many cases where the outer wall heat transfer coefficient is relatively low. This extra surface is effective for most organic substances including hydrocarbons in reducing the size of condenser needed for a specified duty. Water condensing with hydrocarbons creates a new complication

Colonel George A. Burrell (left) admires the Hanlon Award which has just been presented by C. R. Williams, NGAA president.

"INFORMATION PLEASE" session panel of experts in action, reading left to right, J. R. Butler, of J. S. Abercrombie, Inc.; J. G. De Flon of the Flour Corporation; J. B. Zachry, Worthington Pump & Machinery Company; F. J. Colton, Tide Water Associated Oil Company; J. E. Bludworth, consultant; Max Landry, Bruce-McBeth Engine Company; O. L. Smith, The Happy Company; G. L. Brennan, Warren Petroleum Corporation; Harold M. Smith, U. S. Bureau of Mines; W. W. Scheumann, Cities Service Oil Company; R. N. Williams, Elliot Company; Ray Carter, Stanolind Oil & Gas Company; and Merle Church, Shell Oil Company.

due to the presence of two liquid phases in contact with the fins.

Laboratory measurements were reported of heat transfer coefficients attained while condensing eutectic binary mixtures of water with pentane, isooctane, benzene, and chloro-benzene. Water in low concentrations decreases the heat transfer coefficients while at concentrations of 30 percent or more the coefficients are above those for hydrocarbons. The decrease in heat transfer is due to the water accumulation between the fins.

"A Study of Cycling Operations in the West Austin Sands" by Charles B. Johnson, Jr., Chief Engineer, The Chicago Corporation, Corpus Christi, Texas, was presented by Dr. Frank H. Dotterweich, director, Division of Engineering, Texas College of Arts and Industries, Kingsville, with a descriptive commentary by the author.

A discussion providing information on the use of the super-expansibility factors in high pressure gas measurement was presented by E. P. Valby, gas engineer of the Richfield Oil Corporation. Deviations of natural gases and other gases from the perfect gas law have been recognized for many years. The expanding use of orifice meters and various texts made to determine orifice meter coefficients, emphasized the need for deviation factors for the particular gases being measured. A deviation test apparatus was developed by H. S. Bean of the United States Bureau of Standards, and the use of this apparatus and its subsequent modifications has made available a source of experimental data on the behavior of a variety of hydrocarbon gases and mixtures.

Economics Session

Albert H. Weil, United Gas Pipeline Company, was chairman of an economics session which considered the significance of liquids from natural gas, the maximum recovery of hydrocarbons by conservation measures, and process methods to improve LPG recovery in gasoline plants.

NGAA president, C. R. Williams of The Chicago Corporation, Corpus Christi, discussed the "Significance of Liquids from Natural Gas."

"The history of the natural gasoline industry is such that there has never been a marker established with which succeeding periods of production can be compared to determine whether or not a so-called record may have been set or passed. The industry has never decelerated for a sufficiently long period of time to determine such values," he said. "Since the inception of the cycling plant, the direct sale of products from gas has become a first class competitor of the refineries for a share of the expanding domestic motor fuel market."

Domestic users of liquid petroleum gases in 1947 consumed 41 percent of the total production of the natural gasoline industry. Production in 1947 was 130 million barrels and represented an increase of 16.3 million barrels over 1946, he pointed out.

"The recovery of these products is really a huge conservation measure," Williams reminded, "resulting in the salvaging of materials that otherwise would not have been recovered.

Texas Railroad Commissioner William J. Murray, Jr., in presenting "Maximum Recovery of Hydrocarbons," urged that every effort be made to recover the maximum hydrocarbon constituents from oil and gas reserves.

"Such a course," Murray declared, "will prolong the life of all reserves, increase profits of the industry and tend to prevent nationalization of the oil industry. Both oil producer and gasoline plant operator have a responsibility to the public, and if the profit motives and voluntary cooperation do not work to attain the maximum possible recovery of all petroleum hydrocarbons, then the public is liable to step in, assert that it has an interest, and demand compulsory conservation measures."

"Modern Process Methods to Improve LPG Recovery" by G. W. McCullough, K. H. Hachmuth, and A. J. Miller, Phillips Petroleum Company, explains how knowledge of absorption variables, gained during the previous years, now allows the selection of optimum operating conditions for economical high propane extraction. (This article is published in this issue of PETROLEUM REFINER.)

A number of experiments demonstrating the "Inflammability of Gases" were presented by G. M. Kintz and H. F. Browne, U. S. Bureau of Mines, Dallas. Adequate safety education in the hazardous characteristics of liquid hydrocarbons and gases is a necessary step in the prevention of serious plant accidents. Kintz demonstrated conditions under which such accidents could occur and urged that an adequate safety education for plant workers be provided.

"More Propane for Less Money" by Z. C. Ambrose, Southwest Gas Producing Company, Inc., described a process scheme designed for 80 percent propane extraction and includes dehydration of wet gas charge followed by refrigeration before absorption; use of a low molecular weight absorption oil, made possible by low absorber temperature; and "dry" distillation of rich oil in a two-stage stripping system. Use of a low molecular weight absorption oil permits circulation of about one half as much absorption oil as would be required in an equivalent plant of conventional design. The plant, through a two-stage stripping system, avoids use of direct fired heaters for stripping the rich oil, even though the heavier products have a boiling range in excess of those obtained in many high pressure gasoline plants.

Absorber operation at lowered temperature provides some increase in desirable selectivity, thus, eliminating much of the troublesome fixed gases which are normally picked up.

Industry Status

R. M. Chan, vice president of Magnolia Petroleum Company, in "The Oil Industry Today" reviewed present conditions, outlined the supply and demand picture during the past several years, discussed factors influencing the present rising demands, and suggested measures that might help in educating the public to practice oil conservation.

During the motor fuels session, James E. Pew, Sun Oil Company, chairman, discussed on the efficiency of fuels and engines, and the utilization of alcohol-water injection to increase the efficiency of engines were presented.

A reversal of the conditions which have caused the steady disappearance of private risk capital from the American scene will be necessary if the $12 to $15 billion the oil industry will need for expanded operations in the next ten years are to be available, according to Richard Wagner, president of The Chicago Corporation, in "Risk Capital in the Petroleum Industry."

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Temperature control demands

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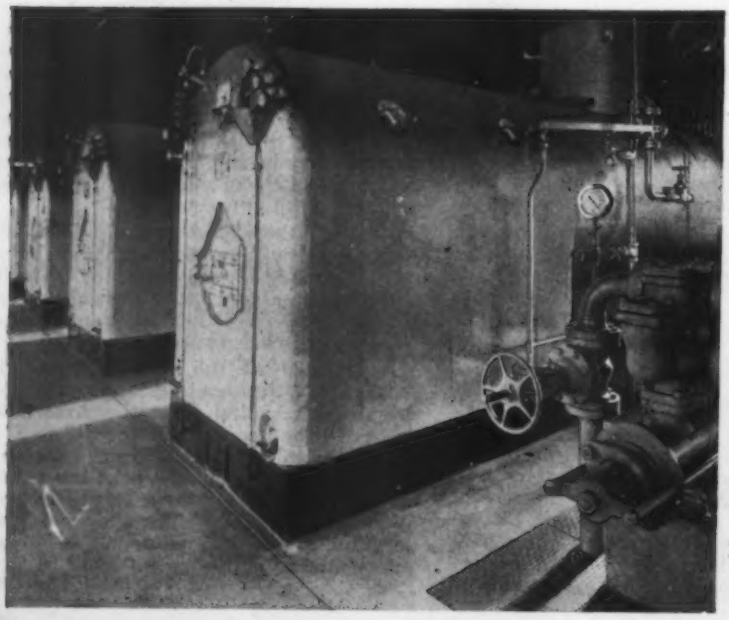

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U. S. Oil and Gas Reserves

Estimates by API and AGA disclose figures unsatisfactory in relation to production

NEW reserves of crude oil during 1947 somewhat more than compensated for production during the year, but the 2.9 percent net gain in reserves was much less than the 7.2 percent increase in the annual rate of production, the API report released last month disclosed. While U. S. crude reserves of 21,487,685,000 barrels at the end of 1947 represented a new all-time peak, they were equivalent to only 11.6 years' supply on the basis of the production of 1,850,445,000 barrels in 1947. The reserves of 20,873,560,000 barrels at the end of 1946 had been equivalent to 12.1 years' supply on the basis of the production of 1,726,348,000 barrels in 1946.

The API estimated crude reserves and the AGA natural gas, and they collaborated on the estimates of natural gas liquids.

This dwindling of years' supply of crude in reserve, in the face of exceptionally active drilling did not mean a great deal.

In the three decades since 1918, proved U. S. reserves have sometimes been as much as 16 to 17 times the prevailing annual production. With the exception of several years in the middle 1920's, they have never been less than 12 times prevailing annual production.

Under the reasonable assumption that U. S. crude reserves should have been maintained in 1947 at least at the end-of-1946 ratio of 12.1 times annual production, it is indicated that at the end of 1947 the reserves should have been at least 22,390,384,500 barrels. Instead, they were actually estimated at only 21,487,685,000 barrels. This indicated a deficiency of 902,699,500 barrels in reserve finding in 1947.

Instead of proving up the estimated total of 2,464,570,000 barrels of new reserves in 1947, it would have been necessary to block out the 902,699,500 barrels additional, or a total of 3,367,269,500 barrels of new reserves to prevent deterioration of the reserves position. The proving up of reserves would have had to be 36.6 percent more than the actual performance.

New Estimates

The new estimates of API and AGA indicate that the total proved liquid hydrocarbon reserves in the U. S. as of December 31, 1947, amounted to 24,741,660,000 barrels. That was 704,881,000 barrels more than the corresponding reserves of 24,036,779,000 barrels at the end of 1946. These figures include natural gas liquids as well as crude oil. Crude reserves were estimated at 21,487,685,000 barrels at the end of 1947, up 614,125,000 from the 20,873,560,000 barrels estimated at the end of 1946. Reserves of natural gas liquids, including condensate, natural gasoline, and liquefied petroleum gases, were estimated at 3,253,975,000 barrels December 31, 1947, up 90,756,000 from the 3,163,219,000 barrels at the end of 1946.

The estimates include only oil and natural gas liquids recoverable under existing economic and operating conditions. The estimations do not include crude oil in the unproved portions of partially developed fields; oil in untested prospects; oil that may be present in regions believed to be favorable; oil that becomes available by secondary-recovery methods; liquid fuels that may become available through chemical processing of natural gas; or synthetic fuels that may be made from coal, shale, or other materials.

Gas Reserves

IN SPITE of comparatively high rates of drilling in 1947, the proving up of reserves of natural gas and natural gas liquids in the U. S. was not sufficiently successful to prevent declines during the year in reserves of both resources in terms of years' supply at prevailing rates

Oil Reserves and Production of the United States

All Figures in Barrels

YEAR	Estimated Proved Reserves Beginning of Year	*New Reserves Proved During Year	Crude Oil Production During Year	Estimated Proved Reserves at End of Year	Ratio of Reserves At End of Year to Year's Production (Year's Supply)	Net Increase or Decrease in Reserves During Year
Reserves Including Condensate						
1918	5,900,000,000	655,928,000	355,928,000	6,200,000,000	17.4	+ 300,000,000
1919	6,200,000,000	878,367,000	378,367,000	6,700,000,000	17.7	+ 500,000,000
1920	6,700,000,000	942,929,000	442,929,000	7,200,000,000	16.3	+ 500,000,000
1921	7,200,000,000	1,072,183,000	472,183,000	7,800,000,000	16.5	+ 600,000,000
1922	7,800,000,000	357,531,000	557,531,000	7,600,000,000	13.6	— 200,000,000
1923	7,600,000,000	732,407,000	732,407,000	7,600,000,000	10.4	
1924	7,600,000,000	613,940,000	713,940,000	7,500,000,000	10.5	— 100,000,000
1925	7,500,000,000	1,763,743,000	763,743,000	8,500,000,000	11.1	+1,000,000,000
1926	8,500,000,000	1,070,874,000	770,874,000	8,800,000,000	11.4	+ 300,000,000
1927	8,800,000,000	2,601,129,000	901,129,000	10,500,000,000	11.7	+1,700,000,000
1928	10,500,000,000	1,401,474,000	901,474,000	11,000,000,000	12.2	+ 500,000,000
1929	11,000,000,000	3,207,323,000	1,007,323,000	13,200,000,000	13.1	+2,200,000,000
1930	13,200,000,000	1,298,011,000	898,011,000	13,600,000,000	15.1	+ 400,000,000
1931	13,600,000,000	251,081,000	851,081,000	13,000,000,000	15.3	— 600,000,000
1932	13,000,000,000	85,159,000	785,159,000	12,300,000,000	15.7	— 700,000,000
1933	12,300,000,000	605,656,000	905,656,000	12,000,000,000	13.3	— 300,000,000
1934	12,000,000,000	1,085,065,000	908,065,000	12,177,000,000	13.4	+ 177,000,000
1935	12,177,000,000	1,219,596,000	996,596,000	12,400,000,000	12.4	+ 223,000,000
1936	12,400,000,000	1,763,087,000	1,099,687,000	13,063,400,000	11.9	+ 663,400,000
1937	13,063,400,000	3,723,028,000	1,279,160,000	15,507,268,000	12.1	+2,443,868,000
1938	15,507,268,000	3,055,233,000	1,214,355,000	17,348,146,000	14.3	+1,840,878,000
1939	17,348,146,000	2,399,828,000	1,264,962,000	18,483,012,000	14.6	+1,134,866,000
1940	18,483,012,000	1,894,717,000	1,353,214,000	19,024,515,000	14.1	+ 541,503,000
1941	19,024,515,000	1,967,009,000	1,402,228,000	19,589,296,000	14.0	+ 564,781,000
1942	19,589,296,000	1,880,142,000	1,386,645,000	20,082,793,000	14.5	+ 493,497,000
1943	20,082,793,000	1,486,972,000	1,505,613,000	20,064,152,000	13.3	— 18,641,000
1944	20,064,152,000	2,066,983,000	1,677,904,000	20,453,231,000	12.2	+ 389,079,000
1945	20,453,231,000	2,087,237,000	1,713,655,000	20,826,813,000	12.2	+ 614,125,000
Reserves of Crude Oil Only						
1945	19,784,530,000	1,870,971,000	1,713,655,000	19,941,846,000	11.6	+ 157,316,000
1946	19,941,846,000	2,665,653,000	1,733,939,000	20,873,560,000	12.0	+ 931,714,000
1947	20,873,560,000	2,470,232,000	1,856,107,000	21,487,685,000	11.6	+ 614,125,000

* In both new and older reservoirs and reflecting upward or downward revisions of earlier estimates based on additional information.
Source: Reserves from American Petroleum Institute and Production from Bureau of Mines.

Natural Gas Reserves and Consumption of the United States

YEAR	Reserves at End of Year (Billions of Cubic Feet)	Authority for Estimate of Reserves	Consumption During Year (Marketed Production: Billions of Cubic Feet)	Years' Supply in Reserve
1918...	15,000	E. Holley Poe et al in report	721	20.8
1919...	15,000	September 22, 1947	746	20.1
1920...	15,000	" " "	798	18.8
1921...	15,000	" " "	662	22.7
1922...	15,000	" " "	763	19.7
1923...	15,000	" " "	1,007	14.9
1924...	15,000	" " "	1,142	13.1
1925...	23,000	*Gas Facts, 1945 and 1946*	1,189	19.3
1926...	23,000	" "	1,313	17.5
1927...	23,000	" "	1,445	15.9
1928...	23,000	" "	1,568	14.7
1929...	23,000	" "	1,918	12.0
1930...	46,000	Earl P. Hindes, then Cities Service Co.	1,943	23.7
1931...	46,000	" " "	1,686	27.3
1932...	46,000	" " "	1,556	29.6
1933...	46,000	" " "	1,555	29.6
1934..	62,000	Ralph E. Davis, Consulting Engineer	1,771	35.0
1935..	62,000	" " "	1,917	32.3
1936..	62,000	" " "	2,168	28.6
1937..	66,000	" " "	3,408	19.4
1938..	70,000	Lyon F. Terry, Chase Nat'l. Bank	2,296	30.5
1939..	70,000	" " "	2,477	28.3
1940..	85,000	Ralph E. Davis, Consulting Engineer	2,660	32.0
1941..	113,800	Petroleum Administration for War	2,813	40.5
1942..	110,000	" " "	3,053	36.0
1943..	110,000	" " "	3,415	32.2
1944..	133,500	" " "	3,711	36.0
1945..	147,789	American Gas Association	3,919	37.7
1946..	160,576	" "	4,040	39.7
1947..	165,927	" "	4,400	37.7

Summary of U. S. Reserves of Natural Gas Liquids, Including Condensate, Natural Gasoline, and Liquefied Petroleum Gases

MILLIONS OF BARRELS

Total proved reserves as of December 31, 1946	3,163.2
Extensions, and revisions of previous estimates,	192.2
New reserves discovered in 1947	59.3
Total proved reserves added in 1947	251.5
Total proved reserves as of December 31, 1946, and new proved reserves added in 1947	3,414.7
Deduct production during 1947	160.8
Total proved reserves of natural gas liquids as of December 31, 1947	3,253.9

of production, the American Gas Association's report last month showed.

In terms of actual cubic feet and barrels, there were small net increases in these reserves, after figuring in the year's production. But these were far too small to leave as many years' supply of reserves at the end of the year as at the beginning. For production of both natural gas and natural gas liquids was sharply higher in 1947 than in 1946, and without reduction in years' supply of reserves, the reserves would have had to be increased much more than they actually were.

Indicated declines in years' supply of natural gas and natural gas liquids in reserves are not matters for alarm. No early shortages of these resources are indicated, for in spite of only small increases during the year, the reserves still are comparatively large. Both are still relatively more plentiful than crude.

Estimates of natural gas reserves by the American Gas Association, showed that the total reserves of 165,926,914,000,000 cubic feet as of December 31, 1947, were equivalent to 29.5 years' supply on the basis of the net production of 5,629,811,000,000 cubic feet in 1947. (Net production indicates gross withdrawals less gas reinjected into the underground reservoirs.) While this was still a comfortable ratio, it was considerably lower than that of a year previously. The estimated reserves of 160,575,901,000,000 cubic feet as of December 31, 1946, were equivalent to 32.5 years' supply on the basis of the net production of 4,942,617,000,000 cubic feet in 1946.

This sharp drop in years' supply of natural gas reserves was primarily the result of an extraordinary increase in demand and production, occurring while reserves were being expanded only moderately or normally. The net production of 1947 was 13.9 percent greater than that of 1946. Instead of increasing 13.9 percent in 1947 as did production, the proved natural gas reserves showed a net increase, after production, of only 3.3 percent.

The present 29.5 years' supply of natural gas is large in comparison with crude oil reserves, which have ranged over many years from ten to 20 years' supply and which have been around 12 years' supply in recent years.

Nevertheless, the current trend toward fewer years' supply of natural gas in reserve has some highly significant implications. In the first place, a continuing improvement in the market for natural gas is indicated. Another obvious implication is that natural gas reserves do not justify too much enthusiasm over currently proposed synthetic plants for making natural gas into liquid fuels.

The year's record as regards reserves of natural gas liquids, as estimated jointly by the American Gas Association and the API, includes condensate, natural gasoline, and liquefied petroleum gases. The estimated reserves of 3,253,975,000 barrels as of December 31, 1947, were equivalent to 20.2 years' supply on the basis of the net production of 160,782,000 barrels in 1947. The estimated reserves of 3,353,887,000 barrels as of December 31, 1946, represented 25.9 years' supply on the basis of the net production of 129,262,000 barrels in 1946. The key to this sharp drop in years' supply is the phenomenal increase in production, which was 24.4 percent greater in 1947 than in 1946. On this basis, reserves would have to be doubled within four years in addition to supplying current needs in order to prevent decline in reserves in terms of years' supply.

Proved reserves, AGA explained, may be in either the drilled or undrilled portion of a given field. Where the undrilled areas are considered proved, they are so related to the developed acreage and to the known field geology and structure that their productive ability is considered assured. Recoverable reserves of natural gas are defined by AGA as the reserves estimated to be producible under present operating practices. Since the estimates are made by fields, the recovery factors or abandonment pressures used in the calculations were determined by the operating conditions of each individual field. Proved recoverable reserves of natural gas liquids are those contained in recoverable gas reserves and extractable by methods now used in the respective fields.

Trends of Operations and Changes in Stocks

Figures on crude stocks are from Bureau of Mines weekly reports; all others from American Petroleum Institute weekly reports, which are estimates on Bureau of Mines basis.

(All figures in thousands of barrels—add 000)

Trends in Week Ended	Crude Oil: Production Daily	Crude Oil: Runs to Stills Daily	Crude Oil: Stocks Week End	Gasoline: Production Weekly	Gasoline: Stocks Week End	Gasoil and Distillate: Production Weekly	Gasoil and Distillate: Stocks Week End	Residual Fuel: Production Weekly	Residual Fuel: Stocks Week End
1947:									
January 25	4,672	4,820	221,655	14,624	99,801	5,630	50,257	8,224	48,558
February 22	4,786	4,860	224,580	14,668	103,904	5,929	40,739	9,532	44,919
March 29	4,865	4,843	225,720	14,396	107,576	5,969	32,737	8,668	43,364
April 26	4,930	4,725	234,051	14,213	103,860	5,435	32,286	8,186	42,668
May 31	5,024	5,000	239,370	14,709	95,867	5,732	36,032	8,910	45,224
June 28	5,109	5,093	236,221	16,070	91,806	5,606	41,721	8,298	48,490
July 26	5,084	5,162	231,357	16,142	85,812	5,708	47,097	8,738	52,497
August 30	5,157	5,209	228,281	16,753	84,105	5,919	54,808	8,962	55,242
September 27	5,196	5,290	224,994	16,501	82,472	6,136	60,406	8,580	56,777
October 25	5,295	5,165	224,308	16,529	82,003	6,638	62,609	8,666	57,419
November 29	5,257	5,309	224,540	16,667	86,319	6,382	61,988	8,405	56,747
December 27	5,285	5,543	222,770	17,013	91,269	7,041	50,324	9,698	51,896
1948:									
January 31	5,318	5,434	221,663	15,986	102,973	7,690	40,580	9,141	50,839
February 28	5,387	5,463	220,507	15,796	111,040	7,963	33,836	8,824	49,206
March 27	5,377	5,324	*220,971	15,783	112,991	7,682	30,717	8,854	48,334

* March 20.

Supply of Oil for U. S. Consumption To Be Studied by API Subcommittees

The petroleum industry has launched a world-wide investigation into the availability of petroleum products for consumption in the United States for both immediate and future needs, President William R. Boyd, Jr., of the American Petroleum Institute announced last month.

The sub-committees of the Institute's National Oil Policy Committee will make the studies, one headed by Howard W. Page, Standard Oil Company (New Jersey), New York, relating to next winter's supply and the second group headed by L. F. McCollum, Continental Oil Company, Ponca City, Okla., to study supply on a 10 to 15-year range.

In addition to these API committees, appointed by the chairman of the National Oil Policy Committee, A. Jacobsen, Amerada Petroleum Corporation, New York, Boyd said he had invited presidents of six other oil trade associations to designate personnel to participate in the discussions and collaborate in the report of Page's committee and to participate in the discussions and report of McCollum's group.

"The alarmed prediction that 'we are running out of oil' with which this idustry has been confronted since 1860 again is abroad in the land," Boyd said.

"There is no question that the nation's position in regard to the availability of liquid fuels deserves close attention. These committees, composed of the best-informed men in the business, will attempt to evaluate, based upon all data at their command, the exact position of the United States in this regard."

The Page committee report is in the hands of the API national oil policy committee and will be presented to API's executive committee April 15.

Membership of McCollum's "long range" subcommittee:

R. G. Dunlop, Sun Oil Company, Philadelphia; R. G. Follis, Standard Oil Company of California, San Francisco; Jake L. Hamon, Cox & Hamon, Dallas; B. A. Hardey, Shreveport, La.; John M. Lovejoy, Seaboard Oil Company of Delaware, New York; A. C. Mattei, Honolulu Oil Corporation, San Francisco; N. C. McGowen, United Gas Pipe Line Company, Shreveport; R. Ogarrio, The Texas Company, New York; A. W. Peake, Standard Oil Company (Indiana), Chicago; John R. Suman, Standard Oil Company (New Jersey), New York; S. A. Swensrud, Gulf Oil Corporation, Pittsburgh; L. S. Wescoat, The Pure Oil Company, Chicago; H. C. Wiess, Humble Oil & Refining Company, Houston and Ralph T. Zook, The Sloan & Zook Company, Bradford, Pennsylvania.

Membership of Page's subcommittee includes:

John W. Boatwright, Standard Oil Company (Indiana), Chicago; Austin Cadle, Standard Oil Company of California, San Francisco; J. E. Dyer, Sinclair Refining Corporation, New York; R. J. Gonzalez, Humble Oil & Refining Company, Houston; Edward T. Knight, The Atlantic Refining Company, Philadelphia; Albert J. McIntosh, Socony-Vacuum Oil Company, New York; Robert N. Sears, Wasatch Oil Company, Salt Lake City and Fred Van Covern, API, New York.

Presidents of trade association invited to participate in these oil policy discussions are, J. B. Fisher, Bradford Pa., National Petroleum Association; Merle Becker, St. Louis, Independent Petroleum Association of America; F. M. Porter, Oklahoma City, Mid-Continent Oil & Gas Association; Hugh A. Stewart, Denver, Rocky Mountain Oil & Gas Association; Ralph B. Lloyd, Los Angeles, Western Oil & Gas Association; and H. T. Ashton, St. Louis, Western Petroleum Refiners Association.

Sees Middle East Oil as Essential for the U. S.

By 1951-52 the Western Hemisphere's oil surplus will disappear and by 1955 it will be needing to receive half a million barrels daily from the Eastern Hemisphere, mostly the Middle East, S. A. Swensrud, executive vice president of Gulf Oil Corporation said in a speech last month. He estimated that the United States will most likely be receiving by 1951 "more than half a million of barrels daily net imports and by 1955 will probably need upwards of a million barrels daily." About half or perhaps less will come from South America and the rest "will have to come from the Middle East," he said.

Because of staggering costs, Swensrud foresaw little probability that synthetic plants, producing oil from gas, coal or shale, could make any substantial contribution to this million barrels a day shortage by 1955.

Processes for the manufacture of gasoline from natural gas are now being developed, but there is yet no evidence that gasoline can be manufactured from coal or shale on a commercial basis, he said. The plants alone to produce a million barrels daily of synthetic liquid fuels from coal probably would cost upwards of $8 or $10 billion or even more.

On the other hand, Middle East crude can be produced and brought to this country "at an economical cost." Pipe lines and tankers, and the conventional refining capacity required to bring to this country and to process a million barrels daily of Middle East oil would involve a capital outlay in the neighborhood of $2½ billion, and no new technical problems would be involved, he pointed out.

Another major problem of the oil industry, Swensrud said, is that profits are not high enough to take care of huge capital expenditures necessary to supply increasing demands for petroleum products.

The difficulties of maintaining maximum production and the extremely high cost of catalytic cracking units make unlikely any important increase in the octane rating of fuels in the near future, but a saving in gasoline will be made possible eventually by the trend to higher compression and other engine developments, he concluded.

CFR Knock Rating Engine Operators Form New Group

Meetings every three months will be held by the newly organized Ark-La-Tex Section of the Texas Regional Group of CFR Knock Rating Engine operators which was organized in Shreveport late in February. Engine operators from Arkansas, North Louisiana, East Texas and Mississippi are included. A monthly exchange of samples was instituted.

R. F. Galloway, Magnolia Petroleum Company, and Texas Regional CFR-EON group leader, assisted in the group's organization. A. C. Johnson, Talco Asphalt and Refining Company, Mt. Pleasant, Texas, was elected chairman.

The next meeting will be in Mt. Pleasant on May 21.

ACS Petroleum Division To Meet Soon in Chicago

Oil chemists will convene in Chicago the week of April 19 for sessions sponsored by the Petroleum Division of the American Chemical Society at the Society's 113th national meeting. Dr. Gustav Egloff of Chicago is chairman of the division.

Eight papers will be presented at a session on "Modern Motor Fuels" on April 19, these to include "The Evaluation of Additives and Blends for Improving the Cold-Starting Characteristics of Diesel Fuels" by F. M. Nelson and Martin Smolak of Socony-Vacuum Laboratories, Paulsboro, N. J., and "Chemistry of Gum Formation in Cracked Gasoline" by E. L. Walters of Shell Development Company, Emeryville, Calif.

A high polymer forum will be held concurrently in the Palmer House at which developments in plastics, synthetic fibers, rubber and similar materials will be reported.

One section will participate in an all-day symposium on "Gel Formation, Detergency, Emulsification, and Film Formation in Nonaqueous Colloidal Systems."

"The Composition of Shale-Oil Naphtha From Colorado Shale," by John S. Ball, G. U. Dinneen, J. R. Smith, C. W. Bailey and Robin Van Meter of the United States Bureau of Mines Petroleum and Oil-Shale Experiment Station, Laramie, Wyo., will be one of 14 papers read at the general meetings April 21 and 22. "The Origin of Petroleums and Natural Gas in the Light of Recent Research" will be discussed by Benjamin T. Brooks, New York consulting chemist.

Corrosion Chief Topic of Refining Session Program

Corrosion is the subject of top importance on the program of the annual mid-year meeting of the Division of Refining of the American Petroleum Institute at the Benjamin Franklin Hotel, Philadelphia, May 10-13.

The program includes an evening session on Tuesday, May 11, on corrosion problems and another the following morning.

Dr. Robert E. Wilson, Chairman of the Board of the Standard Oil Company (Indiana) and vice president for refining of the Institute, will preside at an open dinner session Wednesday, May 12.

The entire program has been designed to be of benefit to all refinery personnel in supervisory positions rather than to technical men only, it was explained.

"Executives of the nation's refineries

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are asked to release a representative number of their men to attend this meeting," Dr. Wilson said. "The more refiners who hear the program and take part in the discussions, the more understanding we will reach on refinery operating problems."

Other subjects on the program are a discussion of training in refining, a part of the Institute's national program of fostering education for men in the oil industry; problems of the production of butane gas, and of high octane motor fuel.

Warren Petroleum Names Don M. Mattocks Secretary

Don M. Mattocks has been elected secretary of Warren Petroleum Corporation and its subsidiaries, succeeding John T. Oxley, resigned.

Mattocks

In his new capacity, Mattocks will head a department which will correlate many of the activities resulting from the expanded scope of company operations. Since the acquisition by Warren of the Hanlon properties in 1946, he has been first assistant to James E. Allison, vice-president and general counsel.

Mattocks' association with Warren dates back to 1940 when he joined the organization as an attorney. He was previously employed in the legal department of the former Prairie Oil & Gas Company, and when this latter company was merged with the Sinclair interests, he was transferred to the legal department of Sinclair Prairie Oil Company.

Public Relations Program Discussed by Oil Group

The national Oil Industry Information Committee in session at Los Angeles late last month reviewed long and short range objectives of the industry's public relations program. Discussions placed emphasis on the differences in public opinion between those sections which have recently experienced shortage problems in fuel oil and public opinion in the South, Southwest and West Coast areas.

The committee agreed on a program of advertising and published material for distribution through district committees and company channels designed to meet the needs of all areas.

A new public opinion survey taken in February, 1948, by the Opinion Research Corporation revealed that substantial progress had been made in establishing a favorable climate of public understanding in those areas where industry efforts had been pointed at specific problems arising from the fuel oil situation.

The program for the immediate future will be "tailored to reach the general public throughout the nation.

Meetings

April

13-15—Southwestern Gas Measurement Short Course, University of Oklahoma, Norman.
14-16—API, Southwestern Spring Meet, San Antonio, Plaza Hotel.
15—National Petroleum Council, Washington, D. C.
19-20—Industrial Accident Prevention Associations, Annual Convention. Toronto, Royal York Hotel.
19-21—American Society of Lubrication Engineers, annual convention, Buffalo, Statler Hotel.
19-23—ACS, Division of Petroleum Chemistry, Chicago.
21-23—National Petroleum Association, Cleveland, Hotel Cleveland.
26-27—Independent Petroleum Association of America, Wichita, Kans.
27-29—API Lubrication Committee, Tulsa, Mayo Hotel.
28-30—Liquefied Petroleum Gas Association, annual convention, Kansas City, Mo.

May

4- 5—American Gas Association, Natural Gas Department, Houston, Rice Hotel.
4- 6—American Oil Chemists Society, New Orleans, La.
6- 7—API Pacific Coast Spring Meet, Los Angeles, Biltmore Hotel.
7—American Institute of Chemists annual meeting, New York, Waldorf-Astoria Hotel.
9-12—AIChE Regional Meet, Cleveland, Hollenden Hotel.
10-13—API Division of Refining Mid-Year Meet, Philadelphia, Benjamin Franklin Hotel.
10-13—National Fire Protection Association, Washington, D. C., Statler Hotel.
15-22—International Petroleum Exposition, Tulsa.
24-26—American Gas Association, joint production and chemical conference, Asbury Park, N. J., Berkeley-Carteret Hotel.
24-28—API Agricultural Development Committee spring meet, Madison, Wis., Lorraine Hotel.
27-28—Natural Gas & Petroleum Assn. of Canada, Niagara Falls, Ontario, General Brock Hotel.
30—to June 4—ASME semi-annual Meeting, Milwaukee, Wis.
31—Gas Technology Short Course, Texas A. & I. College, Kingsville.
31—to June 6—Liquefied Petroleum Gas Assn. annual convention and trade show, Sacramento, Calif.

June

1- 3—Petroleum Industry Electrical Assn., Dallas.
1- 6—LPG Association, Inc., annual convention and trade show, Sacramento, Calif.
2- 4—Annual Short Course in Gas Technology, Texas A. & I. College, Kingsville.
2- 5—Short Course in LP-gas appliances, University of Tulsa.
6- 8—Petroleum Equipment Suppliers Association, Colorado Springs, Broadmoor Hotel.
7-11—Pittsburgh International Conference on Surface Reactions, Mellon Institute, Pittsburgh.
20-23—American Society of Agricultural Engineers, Portland, Ore., Multnomah Hotel.
21-25—American Society for Testing Materials, annual meeting, Detroit, Book-Cadillac Hotel.
30—to July 2—API, Eastern Spring Meet, White Sulphur Springs, Va., Greenbrier Hotel.

September

13-17—Third Instrument Conference and Exhibit, Philadelphia, Convention Hall.
15-17—National Petroleum Assn. meeting, Atlantic City, Hotel Traymore.

October

4- 9—American Gas Association convention, Atlantic City, Ambassador and other hotels.
11-13—National Lubricating Grease Institute, annual convention, Chicago, Edgewater Beach Hotel.

November

8-11—American Petroleum Institute annual meeting, Chicago, Stevens Hotel.
29—to Dec. 4—18th National Exposition of Power and Mechanical Engineering, New York, Grand Central Palace.

Pyrene Recalls 500,000 Suspected Extinguishers

Some 500,000 hand fire extinguishers have been recalled to the factory by Pyrene Manufacturing Company. The company said an "elusive and persistent corrosive" has apparently damaged the working mechanism of many extinguishers and, in time, would probably affect the others listed by Pyrene.

Most of the affected extinguishers are owned by industrial and commercial groups, George Boucher, vice president of Pyrene, said.

According to Boucher, the chemical company supplying the extinguishing fluid to Pyrene has already corrected the defect. He said a national advertising campaign will recall by serial number all of the suspected extinguishers.

Only those hand-pump types manufactured or refilled in the past year and a half are affected, Boucher said. All are 1 or 1½-quart size.

NGAA Sponsors Natural Gasoline Textbook Series

The first section of "Natural Gasoline and the Volatile Hydrocarbons," a new publication, is off the press, according to its sponsor, the Natural Gasoline Association of America, 422 Kennedy Bldg., Tulsa, Oklahoma. The first sections sells for $4 per copy.

Authors of the textbooks, which will be published in five sections, are G. G. Brown, chairman, department of chemical and metallurgical engineering and Donald L. Katz, professor of chemical engineering, University of Michigan; George G. Oberfell, vice president, and Richard C. Alden, director of research, Phillips Petroleum Company. Three of the authors of the book have been presented with the Hanlon Award by the NGAA in past years.

According to the NGAA the new series will constitute a revision of Oberfell and Alden's book "Natural Gasoline," for many years the technical "bible" of the industry.

Printed on pages 8½ by 11 inches, Section 1 has a total of 73 charts and graphs, 36 of which occupy whole pages. Such large scale presentation permits inclusion of data necessarily omitted in smaller reproduction and makes them ready for immediate use. Chapter headings are: "Behavior of Petroleum Hydrocarbons," "Fluid Densities," "Vapor Pressure," and "Critical Properties."

The remaining four sections are in preparation and may be available at approximately eight-month intervals. Section 2's chapters cover vapor liquid equilibria, water-hydrocarbon phase relations, thermal properties, and viscosity and surface tension.

Chapters in Section 3 are on fractionation of hydrocarbons and absorption, adsorption, and distillation. Section 4 is devoted to fluid flow and metering and reservoir behavior and reserves (crude, natural gas and condensate). Chapters XIII and XIV of Section 5 present natural gasoline and condensate operations and utilization of volatile hydrocarbons.

Section 1 gathers in convenient, usable form much of the essential data required for design engineers' calculations, and the bibliography supplies references to other authoritative source data. The plant chemist and engineer will find the work invaluable in checking operations.

Although the book was written pri-

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marily for the technologist, students and others not familiar with involved calculations of petroleum processing will find helpful the many illustrations of data use, solutions of hypothetical problems, and the detailed methods presented for applying the fundamental information in specific calculations. These last named features are expected to give this publication a wide appeal to industry personnel interested in the behavior and processing of hydrocarbons.

Oil Industry Needs for Steel Estimated by NPC

A report estimating direct and indirect steel requirements of the oil industry for an 18-month period starting April 1 at 15,571,500 tons, was compiled last month by the National Petroleum Council.

It now appears that several weeks will elapse before the Commerce Department proposes to the steel industry that it arrange to allocate any part of this total to either oil industry or its supporting industries, if, indeed, any consideration is ever given to the latter.

The Interior Department at the close of March had yet to screen estimates and relay the information to the Commerce Department, but officials of the latter agency indicated industry will have to supply specific evidence to support certain stated requirements or leave the Commerce Department to decide arbitrarily whether these are critical or not.

In particular, criticism was directed at NPC because estimates were submitted in broad groupings by industry branches rather than in terms of actual projects.

The total requirements submitted for the refining branch of the industry was 1,363,800 tons for the 18-month period, or 909,100 tons per year.

Spot Gasoline Shortages Foreseen During Summer

General agreement on the fear of spot shortages resulting from heavy demands in the available supply of gasoline in certain farm districts this summer, and a desire for joint cooperation between farm and oil industries was expressed in Kansas City last month when leaders in those two fields conferred. The meeting was sponsored by the Oil Industry Information Committee.

It was agreed that the job of the farmer is to keep mechanized equipment in good repair and to keep his supplier advised in advance of his needs. Coincidently, the oil men were requested to continue their efforts in delivering the gasoline to the farmer as needed. Leaders urged closest cooperation among oil men and farmers.

CNGA Sets Meeting Dates

The California Natural Gasoline Association will hold its annual June Frolic at the Hondo Golf Club, Downey, Calif., on June 5. An all day sports program, dinner and evening show is planned.

CNGA annual fall meeting will be held Friday, October 8, at the Ambassador Hotel, Los Angeles.

Synthetic Liquid Fuels Program Draws Industry Leaders' Comments

President Truman has signed the case bill (H.R. 2161) into law, so that the Bureau of Mines can continue its synthetic liquid fuels research for three years and authorizes $30 millions in funds.

Secretary of the Interior, J. A. Krug, has submitted a redraft of the Wolverton Bill (H.R. 5475) for synthetic liquid fuels plants, that he says makes clear the government's "intent to promote" the synthetics industry "by industry, rather than by government," The preamble of his bill states that the commercial production of synthetic liquid fuels should be fostered by private industry "as far as practicable," but sets out that government should take such steps as are necessary to "lead the way for industry."

Krug recommended that RFC be authorized to offer inducements, necessary and proper, by way of long term financing, low interest rates, low plant rental, transfer of plant ownership, and subsidies to procure private construction, ownership, and operation of the three plants.

"If RFC Isn't Satisfied—"

Also, the Interior would have authority to name processes, locations, and raw materials. It would draw up long range programs and have $1 million a year for two years for its work in carrying out the bill. No change was suggested concerning the $400 million provided in the Wolverton bill.

But one of the additions in Krug's Bill says that if RFC is not satisfied that private industry will build and operate the three plants proposed, then RFC shall have power to "construct, own, and operate all or any of the plants, itself."

Representative Case has introduced legislation to encourage private investments in synthetic liquid fuel plants by authorizing tax amortization of such investments over a five-year period and permitting 27.5 percent depletion allowance on coal and oil shale used.

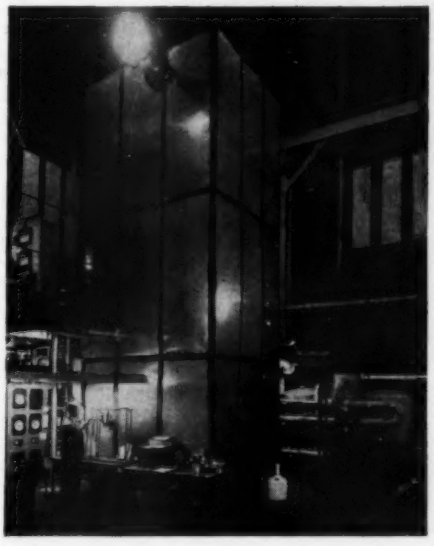
This bill is a companion measure to the amendments to the Wolverton bill, that were made by Senator O'Mahoney, who told the Senate the bill was opening "road to government ownership and operation of a great industry" if private capital is not encouraged to invest in synthetics plants. O'Mahoney's amendments would have allowed tax amortization of private investments in synthetic plants over period of five years instead of present 20-year term and would have permitted 27.5 percent depletion allowance—same as oil—for coal and oil shale used in such plants.

Congress has been asked by President Truman to appropriate $1,500,000 to permit a survey by army engineers to determine sites where commercial synthetic liquid fuels plants can be located. According to the Interior Department, the survey will seek areas where sufficient coal and water supplies are available. Army engineers are doing library work, but won't go into the field until the funds are appropriated.

E. V. Murphree, president of Standard Oil Development Company, proposes virtual elimination of coal hydrogenation from the government's plans for a synthetic fuels industry.

In testimony before the House Commerce committee, he favored a plant for producing oil from shale, but the desirability of two plants for producing oil products from coal, using different processes, he said, is open to considerable question. If it can be indicated from engineering studies that one process for producing oil products from coal is superior to the other, based on present developments, then it is hard to see where anything is gained by erecting a large commercial plant for the less favorable process, he pointed out. Also, Murphree added it appears, on information available, that the coal

New Oxygen Plant



The successful development of a new process for the manufacture of relatively pure gaseous oxygen in tonnage quantities has been announced by the Elliott Company of Jeannette, Pa. A pilot plant, which has been operating for more than two months, "has proven that the Elliott process can operate indefinitely without the necessity of periodic defrosting shutdowns, thus overcoming one of the big disadvantages of similar plants which are unable to stay on stream continuously," the company has announced.

In the process of making oxygen, preconditioned air is cooled to below —300° F., and distilled in a special distillation colmn into its components, oxygen and nitrogen. The oxygen is then piped to the process where it will be used. The nitrogen from the Elliott process "is so pure that it can also be used as a chemical raw material or for bright annealing in steel mills."

Commercial oxygen plants are envisioned which will produce more than 25,000 mcf (1000 tons of oxygen per day. Oxygen on a tonnage basis is a new tool in the steel industry. The Elliott plant "is an American process, designed according to American engineering principles and not based on European practices," the company stated. "It overcomes disadvantages of the Linde-Frankl type process which was developed in Germany and modifications of which are proposed in the United States."

plant should involve a modified Fischer-Tropsch operation rather than dehydrogenation.

P. C. Keith, president of Hydrocarbon Research, Inc., told the House Commerce committee that $1 billion should be appropriated for construction of three commercial synthetic liquid fuels plants instead of the $400 million now proposed by government. Keith said he favored the Wolverton bill with minor modifications.

In addition to suggesting increased appropriation, Keith thought the bill should define only the types of raw materials to be used rather than the processes to be employed to convert them to fuels.

Keith Offers Views

Also, Keith suggested that the size of the commercial plants be 5000 barrels instead of 10,000 barrels because the economic size of any plant is somewhat subject to geographical conditions. Too, he said, the bill should provide for the manufacture of gas from coal to be used as such, because this would reduce demand for heating oil and increase its availability for other fuels.

Eugene Ayres, staff chemist of Gulf Research and Development Company, said he was opposed to passage of the bill because it would make the petroleum supply worse instead of better during the few years immediately ahead and, too, it could have adverse effect on successful long-term development of synthetic fuel projects.

The 300,000 tons of steel for the three plants, he said, could provide for nearly twice as much additional liquid fuel if used to expand domestic petroleum production or to increase domestic refining facilities.

A. C. Rubel, vice president of Union Oil Company of California, before the House Commerce committee described the 120-day death clause in the Wolverton bill as entirely too short. That clause, he said would cause the RFC to proceed immediately to construct one plant of each type, if at the end of 120 days after passage of act private interest had not already contracted to undertake their construction. Rubel suggested that a year should be allowed to permit industry time to carry out preliminary design and engineering work.

Dr. Robert E. Wilson, board chairman, Standard Oil Company (Indiana) wired Secretary Krug urging that a careful study by an industry technical group be made before authorizing a $400 million expenditure by the government on three commercial-size plants.

The benefit of industry technology and considered opinion could not be obtained in a few days he said and "it would take more than a year just to design the proposed plants."

"Even without a study," he said, "I am sure that the proposed building of a hydrogenation plant would be a complete waste of money as both steel requirements and operating costs will be much higher than for Fischer-Tropsch type of operation."

Wilson believed large scale development of coal gasification should be undertaken soon, but conversion of that gas into liquids should wait for experience on large scale synthesis from natural gas, which information will cost the government nothing.

"If Interior Department really wants industry cooperation," he wired, "in construction and operation of these plants, the bill should make some provisions for a plant to be built and operated by a group of companies, and the bill should make some provision to cover losses during operation. Such operating subsidies must be considered in analyzing the projects whether they are built by government or industry."

FPC Engineer's Opinion

William H. Lyon, Federal Power Commission engineer told the House Armed Services Oil committee that conversion of natural gas to liquid fuels is not likely to become a factor of major proportions in meeting the nation's large and expanding liquid fuel requirements.

According to Lyon, one trillion cubic feet of gas, or roughly 25 percent of total estimated consumption last year would be necessary to produce 300,000 barrels per day of liquid fuels and this amount of liquids would be only about 5 percent of a 6 million barrels per day demand for oil which he predicted would be reached shortly.

He said that large-scale synthetic liquid fuel industry must be founded upon coal and oil shale.

Koppers Gets Contract for New Synthetic Fuels Plant

Koppers Company, Inc., of Pittsburgh, on a bid of $4,413,250, has been awarded contract for the design and construction of a gas synthesis (Fischer-Tropsch process) demonstration plant near Louisiana, Mo., to produce oil from coal.

This is the third and last of the demonstration or semi-commercial scale plants planned under the synthetic liquid fuels research and development program of the Bureau of Mines.

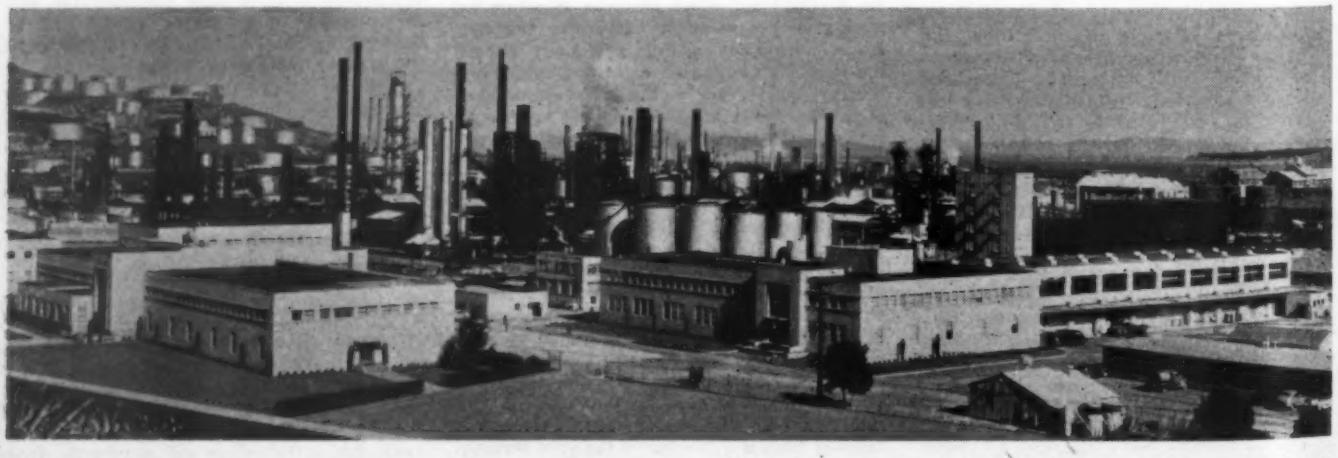
Construction will begin within three months, with the plant to be completed within 15 months thereafter. It will have a capacity of 80 barrels of oil and gasoline daily and will include a coal gasification unit. The gas synthesis plant will be erected adjacent to the Bureau's hydrogenation (Bergius process) demonstration plant under construction near Louisiana, Mo., on the site of the Missouri Ordnance Works, a wartime synthetic ammonia plant. The 200-barrel (daily) hydrogenation plant is expected to be completed next summer.

Although using different processes, both of these plants will employ coal as the raw material for liquid fuels. An oil-shale demonstration plant already is in operation near Rifle, Colo.

The new gas synthesis plant will use many existing facilities and services of the ordnance works which was made available to the Bureau of Mines, Secretary J. A. Krug stated.

Closely synchronized with synthetic fuels research and development labora-

California Research Opens New Research Laboratories



California Research Corporation, a subsidiary of Standard Oil Company of California, last month completed and formally opened its Richmond laboratories, a $2 million postwar expansion and modernization program. Shown above, with the Richmond refinery in the background, the laboratory layout, left to right, is as follows:

Research Laboratory "A," which houses administrative offices and general laboratories; Research Laboratory "C" (one-story building to rear of RL "A"), for industrial lubricant research; Engine Laboratory "A," for testing engine lubricating oils; Process Laboratory "A," which includes cold room and facilities for process research studies using pilot plants; Process Laboratory "B" (seven-story structure in rear of PL "B"), for tall pilot plant operations. Low buildings at extreme right are auxiliary process laboratory units.

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tories at Bruceton, Pa., and Morgantown, W. Va., the coal to oil demonstration plants at Louisiana, Mo., will serve as a proving ground for promising processes developed by Bureau scientists in laboratory and pilot plant, it was explained. As a pattern of scientific, engineering, and economic knowledge is evolved, it will be made available to private industry for projection into plants of commercial size (10,000 barrels or more daily capacity).

The plant site—less than 100 miles above St. Louis on the Mississippi River—is centrally located with respect to the major coal fields of the United States. Coals and lignite from these fields will be tested in the demonstration plants for their processing characteristics and liquid fuel yields.

Of German origin, the gas synthesis or indirect Fischer-Tropsch process to be employed in the new demonstration plant involves first gasifying the coal and then converting the resulting synthesis gas—a mixture of carbon monoxide and hydrogen—to liquid fuels. With American modifications, this process is well adapted to the production of either a good grade of motor gasoline or an excellent Diesel fuel. By-products include wax, alcohols, and hydrocarbon gases. Through cracking, the wax can be converted to Diesel fuel and lubricants, or it may be used directly in polishes, insulating materials, and chemicals. The alcohols also are useful in the chemical industries.

Incorporating the advances made in the Bureau's broad program of research on the Fischer-Tropsch process, the demonstration plant will consist of four distinct parts: 1) coal gasification; 2) gas purification; 3) hydrocarbon synthesis; and 4) refining of products.

Solliday Challenges Cost Figure for Synthetic Fuel

Speaking before the Mid-Continent division of the American Petroleum Institute recently, A. L. Solliday, executive vice-president of Stanolind Oil and Gas Company, challenged statements by the Interior Department that a 2-billion barrels daily synthetic gasoline industry could be built for $9 billion and said he believes such a program would cost $20 to $30 billion instead.

Although he criticized the government's spending of tax money to build commercial - scale synthetic gasoline plants he said that the research program on mining oil shale, producing and refining oil from shale, and coal gasification is valuable and should be continued.

Solliday said that many people are overestimating the number of synthesis plants that can be built using natural gas as the raw material and indicated that probably not more than 20 plants making not more than 150,000 barrels of gasoline daily will be built.

German Fischer-Tropsch technique plays little or no part in the process which will be used in Stanolind's Garden City, Kan., plant, Solliday declared. Rather the process was developed by the American petroleum industry with Stanolind's laboratories making significant contributions.

Regardless of the future prospects in sofar as synthesis techniques are concerned, Solliday stressed that crude oil is in general still the most economic source of liquid fuels and that a healthy crude oil industry will ultimately save billions of dollars in synthetic plant investment and operating costs.

Gen. Hauseman Elected Head of Houdry Process

Brig. Gen. David N. Hauseman, U. S. Army retired, has been elected president of Houdry Process Corporation, Philadelphia, to succeed Eugene J. Houdry, founder of the Company. General Hauseman also will be chairman of the board of Catalytic Construction Company, a Houdry subsidiary.

Hauseman

Under a new policy the board of directors of Houdry Process Corporation will be made up exclusively of men working in the Corporation. The newly-elected directors are: Gen. Hauseman; Arthur V. Danner, executive vice president; Claude C. Peavy, chief engineer; John E. Ford, manager of commercial development; Gordon A. Kessler, manager of patent department; C. G. Kirkbride, manager of laboratories; T. Ellwood Webster, president of Catalytic Construction Company; George H. Daft, comptroller; John D. M. Hamilton, secretary and general counsel.

Eugene J. Houdry, pioneer in the field of catalytic cracking of petroleum products has retired from the company "to devote my time to certain new research problems that are engaging my interest.

"As a shareholder in the corporation, I fully approve this progressive step in turning the management of the company over to the men who are directly carrying on its day-to-day activities," he declared.

Gen. Hauseman has been serving as executive vice president of Temple University, Philadelphia, and president of the school's Research Institute. His headquarters will be in Philadelphia.

Gen. Hauseman has a long military career, serving in both World War I and World War II. He holds the Legion of Merit and the Distinguished Service Medal. He has received both the Army and the Navy Commendation Awards. On his own application, and after 29 years of service, Gen. Hauseman retired from the Army October, 1946.

GE Builds Gas Turbine Power Rail Locomotives

General Electric Company, Schenectady, New York, has announced a new power unit for rail locomotives—the oil-fired gas turbine, which has been operated in factory tests almost continuously since September, 1947. It delivers 4800 horsepower, is 19 feet long, and total weight is 20,000 pounds. Absence of reciprocating parts indicates low maintenance cost. The turbine will be installed on a locomotive chassis shortly.

According to G. W. Wilson, manager of General Electric's locomotive and car equipment divisions, "Initial performance of this experimental unit is promising. However, the potentialities of a new prime mover can be truly evaluated only after exhaustive test in the factory and on the rails."

Mobile Gas Turbine

Mobile gas turbine plants are under development by Allis-Chalmers Manufacturing Company. The proposed units could operate as a sole source of power or could be synchronized with an existing power system.

Simplicity, extremely smooth operation and no requirement for water would characterize the gas turbine plants, according to Allis-Chalmers. Operating on oil, the units would require only fuel line connections to tank cars or storage tanks, in addition to the electric transmission line connection.

The mobile gas turbine plants appear to be a promising source of extra power quickly available. Where regular power plants are out of operation, such units could be located on a rail siding from where they could quickly deliver power into the existing system. Similarly, industrial plants faced with a sudden temporary need for extra power could put a mobile power plant into operation. It could also provide emergency service during service shutdowns or turbine failure.

Plan for Manufacture of New Fuel Solution Given

Thompson Vita-Meter Corporation will market its fuel injection device through automobile jobbers, but manufacture and distribution of the alcohol-water-tetraethyl antiknock solution will be through the oil industry.

C. H. Van Hartesveldt, vice-president, said nothing exclusive will be made in arrangements for making and marketing the solution. Any refiner who will sign an agreement calling for payment of a sales commission to Thompson and assuring manufacture of the solution according to specification will be eligible to enter the field. He said straight run gasoline to be used with the solution should be at least 60 octane.

Would Convert Anthracite Silt Into Usable Gases

A new process is being developed to convert reserves of waste anthracite silt burnable as fuel according to Frank W. Earnest, Jr., president of the Anthracite Institute who said that 200 million tons of this silt are immediately available for such conversion. Average annual anthracite coal production approximates 50 million tons.

Anthracite silt, about the size of granulated sugar, hitherto unusable, has accumulated at the mine heads for years. The conversion process, related to the German method of gasifying brown coal, has been under development at the Institute's laboratories and promises to make feasible the conversion to gas.

Anthracite silt can be processed into three gases, two of which are fuel gases. A pilot plant for converting the anthracite waste into gases is under construction in Wilkes-Barre, Pa. First large-scale use of such gas would be to provide fuel to power anthracite region collieries.

The next logical use of the gas would

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be by petroleum processing companies and by chemical companies. These companies could use the gas to produce synthetic fuel products, such as gasoline, and synthetic chemical products, Earnest said.

Gulf Announces Gasoline Plant for Crane County

Gulf Oil Corporation will construct a natural gasoline plant in Crane County, Texas, approximately 30 miles Southwest of Odessa, Texas, the plant to be a high pressure absorption type gasoline extraction plant. To operate at approximately 600 pounds pressure, it will have a capacity of approximately 40,000 mcf of gas per day.

There will be nine 1200-hp. angle-type gas engine-driven compressors to gather and compress the gas. Four 800-hp. gas gas engine-driven generators will furnish electric power.

The plant which will utilize gas from the Sand Hills, Waddell-Henderson, University Waddell-Ellenburger and C-Bar Silurian Fields, will be designed to extract 80 percent propane, 98 percent butane and all of the 26-pound natural gasoline.

Texaco Lawrenceville Plant to Be Expanded

The Texas Company has announced expansion of its Lawrenceville, Ill., refinery of 20,000 barrels per day to bring the total capacity of the plant to about 50,000 barrels per day.

New facilities will include a 20,000-barrel per day vacuum still, a fluid catalytic cracking unit, and an additional polymerization unit. The gas recovery plant will be expanded and a new high pressure boiler and additional tankage will be installed.

The plant refines West Texas, Illinois, and Midcontinent crude oils.

Monsanto's Rebuilt Plant About Ready for Operation

Monsanto Chemical Company's Texas City plant is expected to be in operation by June. The plant was destroyed a year ago following a ship explosion.

The company's insurance claim of $22 million is in negotiation.

The first unit at Texas City, the ethylene plant, will be running this month.

Tighe Rejoins Chilton

Frank P. Tighe, until recently director of public relations and advertising of Houdry Process Corporation, has returned to Chilton Company as editor of Motor Age. His affiliation with Chilton dates back to 1925 and at one time he was managing editor of Motor Age. Other important posts occupied by Tighe over a period of years include representative of Roche, Williams and Cleary, Inc., general manager of the Philadelphia Automobile Trade Association, and head of the business magazine section of the Office of Censorship, Washington, D. C., during the war.

Cox & Hamon's Gasoline Setup Is Headed by Oxley

John T. Oxley, of Tulsa, has been elected president of the recently organized Texas Natural Gasoline Corporation, an affiliate of Edwin B. Cox & Jake L. Hamon of Dallas. Oxley previously was secretary and assistant treasurer of Warren Petroleum Corporation of Tulsa, and manager of its gas division. He also was secretary of Warren Oil Corporation and Natural Gasoline Corporation and secretary of Warren Maritime Corporation.

Texas Natural Gasoline Corporation will permit Cox & Hamon to become more active in the gas and natural gasoline phases of the petroleum industry.

Oxley worked for Amerada Petroleum Corporation before joining Warren in 1935 at the time that corporation acquired Amerada's natural gasoline department.

Oxley announced appointment of L. A. Boyd, of Tulsa, as chief engineer and general superintendent of the new companies plants. Boyd recently resigned as chief engineer for Warren Petroleum Corporation after having been with the Warren organization since 1940.

Boyd

Boyd has been in the oil business since 1920, serving the old Transcontinental Oil Company, Big Lake Oil Company, Signal Oil & Gas Company, and The Gas & Oil Products, Ltd., Calgary, Alberta, Canada, before joining Warren in 1940. He was in charge of engineering from 1940 to 1948, during which time he supervised the construction of five gasoline plants. He introduced extensive use of air cooling and hydrocarbon vapor stripping at Warren's Hull-Silk plant, eliminating the use of steam.

Claig H. Perry, of Tulsa, has been named secretary and manager of the gas division of the Company. Perry recently resigned from Warren Petroleum Corporation, Tulsa. At the time of his resignation, he was gas engineer in charge of all exploration work in connection with locations of new plants.

Perry's entire business career of 19 years has been spent with two corporations—Amerada and Warren—in the gas-gasoline branches of the petroleum industry.

Refining Capacity Survey Planned by Senate Group

The Senate small business committee will make a new survey of refining capacity in view of a preliminary investigation in the Midwest that showed in some cases as much as 40 percent of independent capacity shut down for lack of crude oil. A survey made last fall showed 113 independent refiners had idle capacity averaging 31.5 percent, while 100 major refiners were operating within 6.4 percent of full capacity.

In the latter part of March, committee staff members made an on-the-spot study, primarily around Omaha, Nebr., on gasoline and tractor fuel outlook for this summer, and reported that even some major refiners are having trouble in getting adequate crude oil supplies.

The new survey will be made to determine if same conditions extend to other sections of U. S.

Indiana Standard Revises Employes' Benefits Plans

A new group hospital expense and surgical operation insurance plan, including physician's hospital attendance benefits, and a revised sickness and disability benefits plans for employes were announced last month by Standard Oil Company (Indiana).

Effective date for the new hospitalization expense and surgical operation insurance plan will depend upon the voluntary enrollment of at least 75 percent of the eligible employes in those groups where the plan is made effective. The revised sickness and disability benefits plan became effective April 1 for more than 20,000 eligible employes throughout 15 north central states.

Cubbage Joins Stanolind

Thomas L. Cubbage has joined Stanolind Oil and Gas Company as operating superintendent in the manufacturing department.

Cubbage, who was formerly plant superintendent of the government-owned, Phillips Petroleum Company-operated Plains butadiene plant at Borger, Texas, will supervise the operations of Stanolind's gasoline and cycling plants and its refinery at Superior, La.

Ole Berg to Toronto

Ole Berg, Jr., of Los Angeles, last month was appointed vice president and assistant to the president of British American Oil, Toronto, Ontario.

Berg comes to the Canadian company from Union Oil Company of California, where he was vice president and manager of general sales.

Pearson and Evans Given Honors at NACE Meet

Dr. John M. Pearson and Dr. Ulick R. Evans have been named recipients of the 1948 awards of the National Association of Corrosion Engineers. Dr.

Pearson

Evans

Pearson has been honored with the Frank Newman Speller Award in Corrosion Engineering and Dr. Evans with the Willis Rodney Whitney Award in Corrosion Science. Presentation was made at the association's annual banquet April 7 at the Hotel Jefferson, St. Louis.

Dr. Evans was honored in recognition of his outstanding contributions to the fundamental knowledge on which the practical control of corrosion is based. Dr. Pearson's award was for his out-

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standing service in the application of scientific research in the prevention of corrosion, particularly by cathodic protection of underground structures.

Dr. Evans has been associated with Cambridge University since 1919. Dr. Pearson is director of the Physical division of the Research and Development department of Sun Oil Company.

Davison Is Promoted By California Standard

W. W. Davison has been appointed a vice president of Standard Oil Company of California. He formerly was general manager of Standard's El Segundo refinery.

Davison

After graduation from the University of California in 1921, Davison joined Standard as a junior engineer. The following year he was made assistant chief engineer and the next year became a designs engineer. In 1928, he was assigned to the manufacturing department, where he served until 1936, when he was made refinery superintendent at Standard Oil Company of British Columbia's plant at Vancouver, B. C.

Four years later he became assistant superintendent at El Segundo and in 1944 was made manager.

Finney Named Head of Salt Lake City Refining

C. E. Finney, Jr., has been advanced to the presidency of Salt Lake Refining Company and Salt Lake Pipe Line Company, subsidiaries of Standard Oil Company of California. In his new position, Finney will direct the construction and operation of both projects. The Salt Lake City refinery will cost $5 million and is scheduled to be completed and in operation by November 1, 1948.

Finney was graduated from Cornell University in 1912 and joined Standard in 1919. He was made assistant to the general manager of the manufacturing department in the home office in 1937 and general manager in 1943. He became a vice president in 1946.

Miller Succeeds Wilson

James A. Wilson, plant manager at the headquarters of Monsanto Chemical Company's Merrimac Division at Everett, Mass., has been appointed production manager for the division.

Wilson is being succeeded as plant manager by Russell L. Miller, plant superintendent.

Bussert Succeeds Fahrney

Standard Oil Company of California has appointed L. M. Bussert as manager of operations in its marketing department. He succeeds P. L. Fahrney, recently elected to a vice-presidency of the company.

Military Petroleum Advisory Committee Names Groups to Study Synthetic Fuels

Bruce K. Brown, vice president of Standard Oil Company (Indiana), Chicago, and chairman of the Military Petroleum Advisory Committee, last month announced members of industry committees who are making a survey of the significance of synthetic fuels in connection with national security for MPAC.

The Synthetic Fuels Committee of MPAC falls in the general jurisdiction of R. G. Follis, Standard Oil Company of California, San Francisco, Calif., who is chairman of the Refining Panel of MPAC. These committeemen, after being nominated by MPAC, were specially appointed for this service by the Army-Navy Petroleum Board and the Department of the Interior.

Here are the committee members:

Synthetic Fuels Committee—A. P. Frame, Cities Service Company, New York, who is also a member of MPAC, chairman; Alois Kremser, Standard Oil Company of California, San Francisco; Max B. Miller, Max B. Miller and Company, New York; E. V. Murphree, Standard Oil Development Company, New York; and J. K. Roberts, Standard Oil Company (Indiana), Chicago.

Subcommittee on Natural Gas—J. K. Roberts, chairman. *Reserves group:* H. D. Hancock, Cities Service Company, New York, group chairman; R. O. Garrett, Arkansas Louisiana Gas Company, Shreveport, La.; K. C. Heald, Gulf Oil Corporation, Pittsburgh; D. P. Olcott, Humble Oil & Refining Company, Houston; T. W. Johnson, Union Producing Company, Shreveport; T. V. Moore, Standard Oil Development Company, New York; and A. M. Ripple, Phillips Petroleum Company, Bartlesville, Okla. *Process group:* L. C. Kemp, Jr., The Texas Company, New York, group chairman; J. H. Arnold, Hydrocarbon Research, Inc., New York; H. Z. Martin, Standard Oil Development Company, Elizabeth (N.J.); J. E. Rouse, Stanolind Oil and Gas Company, Tulsa; and H. H. Storch, United States Bureau of Mines, Pittsburgh.

Subcommittee on Coal—E. V. Murphree, chairman, *Reserves and mining group:* Joseph Pursglove, Jr., Pittsburgh Consolidation Coal Company, Pittsburgh, group chairman; I. N. Bayless, The Union Pacific Coal Company, Omaha, Neb.; J. D. Doherty, United States Bureau of Mines, Washington, D.C.; G. C. Gester, Standard Oil Company of California, San Francisco; and M. H. Forester, Pittsburgh Consolidation Coal Company Pittsburgh. *Process group:* E. J. Gohr, Standard Oil Development Company, New York, group chairman; E. T. Layng, Hydrocarbon Research, Inc., New York; W. C. Schroeder, United States Bureau of Mines, Washington, D.C.; J. E. Seebold, Standard Oil Company (Indiana), Whiting, Ind.; and H. S. Turner, Pittsburgh Consolidation Coal Company, Library, Pa.

Subcommittee on Shale—Alois Kremser, chairman. *Mining and disposal group:* Joseph Pursglove, Jr., group chairman; I. N. Bayless; E. D. Gardner, United States Bureau of Mines, Denver, Colo.; G. O. Gester; and A. C. Rubel, Union Oil Company of California, Los Angeles. *Process group:* Alois Kremser, group chairman; H. G. M. Fischer, Standard Oil Development Company, New York; A. C. Rubel; W. C. Schroeder; and T. P. Simpson (L. P. Evans, alternate), Socony-Vacuum Laboratories, Paulsboro, N.J.

MPAC was created by the Secretary of the Interior at the request of the Secretaries of War and Navy to study the petroleum aspects of military plans and problems and to provide the Army-Navy Petroleum Board and the Oil and Gas Division of the Department of the Interior with counsel, advice, and information.

Short Course Offered In LP-Gas Appliances

The University of Tulsa's short course for LP-gas appliances which will be held June 2-5, will place emphasis on specialized sessions this year. Subjects include hot water heaters, gas refrigeration, liquid meters, fully automatic LP-gas ranges, automatic controls, etc.

Three subjects will be treated in general sessions—customer relations, future of the LP-gas industry, and properties of LP-gas.

Povl Ostergaard Dies

Povl Ostergaard, 48, chief process engineer of Gulf Oil Corporation, died February 17, after a brief illness.

With Gulf since 1926, Ostergaard is principally known for his work in connection with the Gulf polyform process, which he invented and developed. He was either senior author or co-author of numerous technical papers on various phases of that process for the conversion of naphtha or gas oil to gasoline. However, in addition to seven Polyform patents, he also held patents dealing with thermal cracking, fractionation, isomerization, and catalytic cracking. He also was skilled in the mechanical phases of process design.

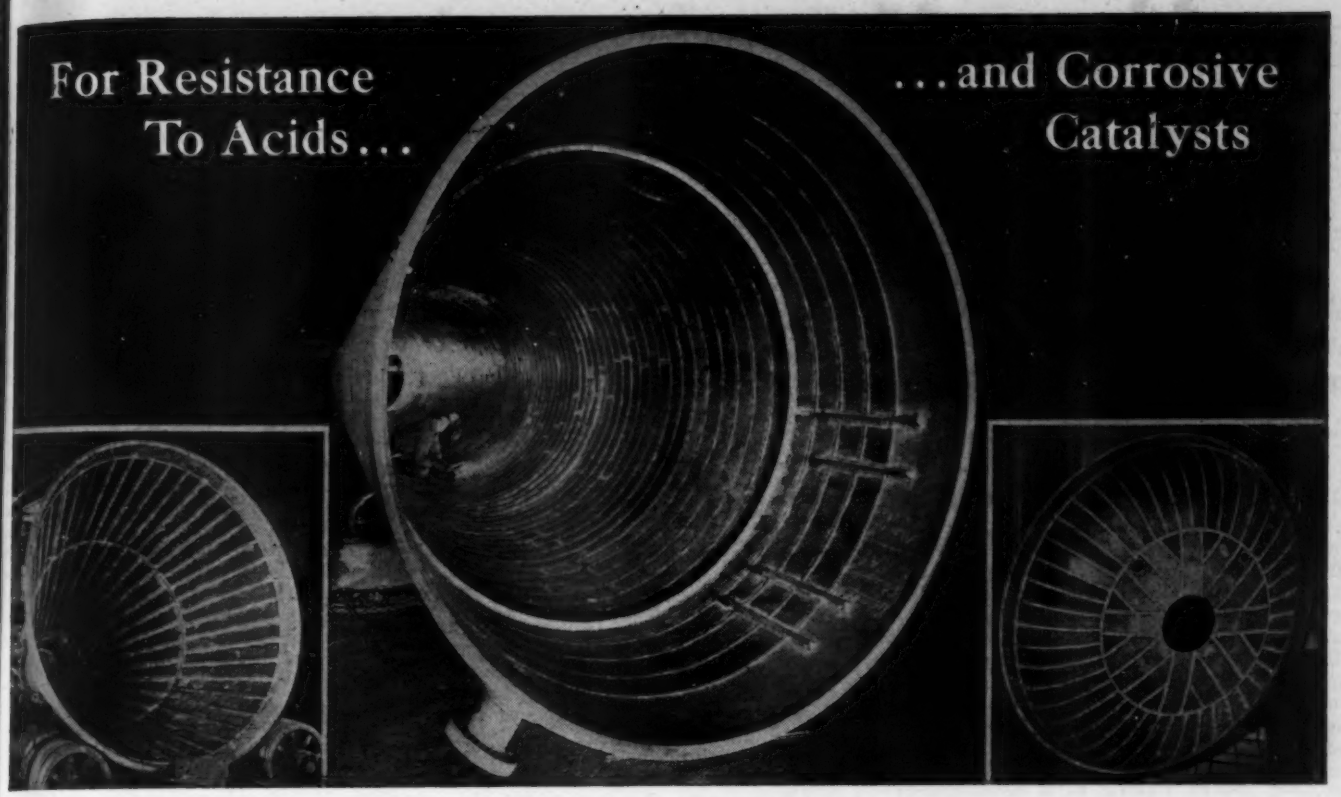
Kelso Leaves Houdry

George Kelso, manager of the technical service and plant operations division of Houdry Process Corporation, Marcus Hook, Pa., resigned on March 15 to become a manufacturer's representative in the Philadelphia area. He will be associated with V. L. Sanderson and their contacts will be with the petroleum and chemical industries.

Prior to and during the war Kelso, who has been with Houdry since 1931, had charge of the initial operation of many catalytic cracking plants.

Texaco Fellowship

The Texas Company is establishing a graduate fellowship at the University of Southern California to encourage advanced studies and research in the field of chemical engineering. The awards will be made by the Dean of the College of Engineering. Its purpose will be to study fundamental problems of distillation in refinery procedure.



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THE MONTH . . .

Burks

Fisher

Esso Standard Directors Get Manufacturing Posts

Dr. H. G. Burks, Jr., a director of Esso Standard Oil Company, has been named general manager of the company's manufacturing operations and H. W. Fisher, also a director, succeeded Dr. Burks as general manager of the East Coast refineries. Both men joined Esso more than 20 years ago upon graduation from the Massachusetts Institute of Technology.

Dr. Burks started as a development engineer at the Bayway refinery in Linden, N. J., and rose through various manufacturing positions to director in 1943. He was appointed manager of East Coast refineries in 1946.

Fisher joined the company as a chemical engineer at the Baton Rouge, La., refinery and much of his career has been devoted to the development and production of chemicals from petroleum. Elected a director in 1945, he later assumed the added responsibility of the presidency of Enjay Company, Inc., a subsidiary of Standard Oil Company (New Jersey) devoted to marketing of petroleum chemical products.

Foster Wheeler Promotions

Martin Frisch, vice president in charge of engineering of Foster Wheeler Corporation, last month was elected a director of the corporation. Frisch joined the company in 1929 and was elected a vice president last year.

Lee Allen Swem, a director of the company, and assistant to the president for the past three years, was elected a vice president. Swem became associated with the company in 1936 as patent counsel.

Stanolind Gift

Stanolind Oil & Gas Company has granted $125,000 toward a $750,000 petroleum science building for the University of Tulsa. Construction of the three-story building to include research laboratories, is scheduled to start before December 1, 1948, under terms of the grant.

Science Bill in Senate

National science foundation legislation, which is expected to comply with White House views, has been handed to the Senate. The bill provides for appointment of the director by president instead of by the foundation board as provided in a bill vetoed last year.

NGAA Convention —

CONTINUED FROM PAGE 169

High taxes, public spending and schemes of economic planners, discouragement by the government false doctrines of a riskless society, false propaganda about business profits and consequent inroads by labor on capital formation, higher costs and inadequate depreciation charges, and inordinate exportation of capital goods needed in the nation's own economy, were listed as the principal reasons for the disappearance of risk capital.

Wagner defined risk capital as "that part of the nation's capital funds which is or may be devoted to finance new products, ideas, methods or services, or to supply equity capital for business expansions."

High Compression Engines

The Research Laboratories Division of General Motors has constructed several six-cylinder engines of 12.5 to 1 compression ratio and J. M. Campbell and L. L. Withrow, of General Motors Corporation, in discussing "Engines and Fuels for Higher Efficiency" presented a comparison of one of these engines with a 1946 production engine of comparable horsepower.

The fuel originally used in the 12.5 compression engine was clear triptane but subsequently the engine has been operated in a car on gasoline having a research octane number in the vicinity of 100 octane plus 0.2 ml. tel. and a Motor octane number between 88 and 90. This experimental engine of 12.5 to 1 compression ratio has demonstrated two important principles: 1) That high compression engines can be made to conform with accepted standards of smoothness and general performance; and 2) that large gains in fuel economy can yet be made over current levels provided suitable fuel becomes available at a reasonable cost to the consumer.

Since 1930 the average efficiency of fuel utilization in passenger car engines expressed in ton-miles per gallon of gasoline, has increased by more than 30 percent, this is the result of the combined efforts of the petroleum and automotive industries.

Looking toward the future, General Motors has run full-scale automobile tests with high compression engines showing that further gains in fuel utilization of the same order of magnitude as those obtained in the past are still possible.

"There is an apparent need for raising the anti-knock quality of gasoline without the large refinery investment and loss of gasoline yield associated with increasing octane numbers," C. H. Van Hartesveldt, vice president, Thompson Viat-Meter Corporation of Cleveland, Ohio, said in presenting "Economics of Anti-Detonant Injection."

"Gasoline engines require high anti-knock quality only when they are at or near full throttle. Most ground vehicles operate under these conditions but a small part of the time, making it economical to supply an anti-detonant only when needed."

Use of Alcohol

"An alcohol-water-tetraethyl lead anti-detonant used through a fully automatic injection device will give octane numbers at costs competitive with refinery methods," Van Hartesveldt continued. "This new source of anti-knock quality can be used in two ways: 1) Added to low octane number, high burning quality gasoline to give adequate octane number and important maintenance savings to commercial fleets; or 2) used with regular and premium grades of gasoline to satisfy the requirements of high compression ratio engines."

In a prepared discussion by R. I. Potter of Standard Oil Company (Ohio) on Van Hartesveldt's paper, Potter compared the performance of conventional engines using regular and premium grades of gasoline with the performance of high compression engines utilizing the injection of an anti-detonant solution to meet peak power demand. The requirements of the base fuel with anti-detonant injection was only slightly higher than the requirements of the standard engine. Further, anti-detonant injection readily allows octane improvement above the octane ratings of present day motor fuels. This is significant because octane numbers at today's higher levels are not as easy to obtain as were an equivalent number of octane points a few years back.

"Any device or method for applying high anti-knock quality to high compression engines during the moments of their peak octane requirements, should be considered by both the automotive and petroleum industries in the interests of conservation of petroleum resources," Potter asserted.

Forum Session

A forum, in which executives and technical men questioned leading authorities of the natural gasoline industry on processing, production, operating trends, and other matters of interest was held as the final session of the convention. Moderator was D. E. Buchanan of the Hiwan Oil and Gas Company of Houston. The panel comprised: For general operations—Ray Carter, Stanolind Oil and Gas Company, Tulsa; J. R. Butler, J. S. Abercrombie, Inc., Houston; F. J. Colton, Tidewater Associated Oil Company, Los Angeles; and Merle Church, Shell Oil Company, Inc., Houston.

For LPG and hydrocarbon chemicals—G. L. Brennan, Warren Petroleum Corporation, Tulsa, and J. E. Bludworth, consultant, Corpus Christi.

Engines and pumps—J. B. Zachry, Worthington Pump and Machinery Company, Tulsa, and Max Landry, Bruce-McBeth Engine Company, Tulsa.

Refining—W. W. Scheumann, Cities Service Oil Company, Lake Charles, La.

Cooling towers and air cooling—J. G. De Flon, Fluor Corporation, Los Angeles, and O. L. Smith, The Happy Company, Tulsa.

Synthetic Hydrocarbons—Harold M. Smith, U. S. Bureau of Mines, Bartlesville, Okla.

Centrifugal blowers—R. N. Williams, Elliott Company, Jeannette Pa.

Question screening committee: James Doughty, Southern Minerals Corporation, Corpus Christi, chairman; W. L. Bowser, The Atlantic Refining Company, Dallas; V. E. Middlebrook, Shell Oil Corporation, Corpus Christi; Frank Stone, La Gloria Corporation, Corpus Christi, and Carl Wharton, Warren Petroleum Corporation, Tulsa.

SCIENCE and TECHNOLOGY ABSTRACTS

The abstracts here presented are selected from the current literature of science and technology to afford reference to fundamental information not easily available to all readers. Abstracts of articles appearing in readily obtainable trade journals are not included.

Photostat copies of original articles will be supplied at cost by The Leslie Laboratories. Complete or limited bibliographies covering special topics by title, by abstracts, or in complete manuscript, will also be prepared and furnished at reasonable cost by the Laboratories.

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Fundamental Physical and Chemical Data

Vapor-Liquid Equilibria in Hydrocarbon Systems—Part II. STUART T. HADDEN, *Chem. Eng. Progress*, 44 (1948) pp. 135-56.

The paper is a continuation of the previous article on the subject in Chemical Engineering Progress in January. The authors present the procedure for obtaining vapor-liquid equilibria in hydrocarbon systems from low pressures up to the convergence pressure of the system. Correlations are presented for obtaining K, equilibrium-vaporization constant, of nonhydrocarbons occurring frequently in hydrocarbon streams either as contaminates or as process materials. Three nomograms are presented and give correlations with adequate accuracy for the experimental ideal values of the equilibrium-vaporization constants of hydrocarbons. Component lines can be constructed for any aliphatic paraffins and mono-olefins. Component lines can be constructed for diolefins and acetylene hydrocarbons but the data from such lines should not be used for mixtures containing paraffins and mono-olefins having boiling points close to those of the diolefins or acetylenes because frequently azeotropic mixtures are formed. Component lines can be constructed for the aromatic and naphthene hydrocarbons but data from such lines should only be used in hydrocarbon mixtures of aromatics or naphthenes. A bibliography of 61 references is included.

Influence of Interatomic Forces on Paraffin Properties. HARRY WIENER, *J. Chem. Phys.*, 15 (1947) p. 766.

Empirical formulas for molecular refractions, molecular volumes, heats of formation and boiling points of the paraffin isomers were reported by Platt. His variables were based on the assumption of additivity of bond properties and of interaction between different bonds in the molecule. The same basic formula is derived by the author without any assumptions as to the nature of the forces interacting between two atoms. Several formulas are compared with values reported in the A.P.I. tables. The accuracy is of the same order of magnitude for molecular refractions, molecular volumes, and boiling points, but the Platt formula is much better for heats of formation. The author's formula employs two empirical parameters, Platt's formula six, and the one by Taylor, Pignocco and Rossini seven.

Dielectric Studies. IX. Studies with Gasoline. NORMAN C. C. LI AND THOMAS D. TERRY, *Jour. Amer. Chem. Soc.*, 70 (1948) pp. 344-5.

The work shows that gasoline and its fractions are essentially non-polar. Dipole moment can be measured using gasoline as a solvent, for it is a material of no definite molecular weight and the physical properties of which vary greatly with different samples. A summary of the dipole moment values obtained using gasoline and similar solvents of indefinite molecular weights is given and the results are shown to be in close agreement with values obtained using non-polar solvents of known molecular weights.

The Use of Tritium in the Determination of the Solubility of Water in Solvents. The Solubility of Water in Benzene. G. G. JORIS AND H. S. TAYLOR, *J. Chem. Phys.*, 16 (1948) pp. 45-51.

Ordinary methods for the determination of water in hydrocarbons require large amounts of the hydrocarbons if accuracy is to be obtained. Water is now available containing radioactive hydrogen. This has made possible the development of a new method for measuring water solubility in hydrocarbons. Radioactive hydrogen emits soft beta-radiations, the intensity of which is best measured by direct introduction of the radioactive sample into the Geiger-Müeller counter. Water dissolved in the hydrocarbon can absorb from the hydrocarbon on calcium oxide by stained reaction. Water in calcium hydroxide can be removed by ethanol vapor if a vacuum is used.

Chemicals Wanted

The National Registry of Rare Chemicals, Armour Research Foundation, 33rd, Federal and Dearborn Streets, Chicago 16, has received urgent requests for the chemicals listed below. If anyone has one or more, even if only one gram quantities, please inform the Registry.

1,16-Hexadecane dicarboxylic acid
1,15-Pentadecane dicarboxylic acid
1,12-Dodecane dicarboxylic acid
1,14-Tetradecane dicarboxylic acid
1,13-Tridecane dicarboxylic acid
Di-n-butylene triamine
Phenylisoxazoles
Benzylisoxazoles
C-Phenylpyrroles
Benzylpyrroles
Diallyldichlorosilane
Methionic acid and diphenyl ester
Chlorogermane
Pyrogallol monomethyl ether monosulfonic acids
Montanic acid
Indole-3-aldehyde
Heneicosane
4-Phenylguaiacol
Catechol monophenyl ether
Catechol monoallyl ether
5-Vinylguaiacol.

Chemical Compositions and Reactions

Aromatics in Petroleum Fractions, Separation and Determination from High Boiling Petroleum Fractions by Adsorption. M. R. LIPKIN, W. A. HOFFECKER, C. C. MARTIN, AND R. E. LEDLEY, *Anal. Chem.*, 20 (1948) pp. 130-134.

A method for the determination of percent aromatics with an accuracy of 1 percent in hydrocarbon solutions boiling above 200° C. is described. The method involved is the dilution of the sample with an equal volume of pentane, followed by percolation through a fixed quantity of silica gel, washing the gel with pentane until the refractive index of the percolate drops to the refractive index of the pentane, and desorption of the aromatics with benzene and methyl alcohol. The method used classifies all hydrocarbons containing one or more aromatic rings as aromatics whether the side chain is paraffin, olefin, naphthene, or cyclo-olefin. A bibliography of 16 references is included.

Pure Compounds from Petroleum, Purification and Purity of Hydrocarbons. FREDERICK D. ROSSINI, *Anal. Chem.*, 20 (1948) pp. 110-121.

American Petroleum Institute Research Project No. 6 started at the Bureau of Standards in 1927 and has been in progress continuously. It has involved a total of approximately 200 man-years of research. A description of this work dealing with the separation of pure compounds from petroleum, and with the purification and determination of purity of hydrocarbons, is presented by the author. This report summarizes the work. The first part of the report surveys the problem of determining the composition of natural petroleum in terms of its individual components. The latter part describes the work of puri-

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fying and determining the purity of hydrocarbons required for use as standard samples for calibrating analytical apparatus and instruments and for measuring needed physical, thermodynamic, and spectrographic properties. Table 1 of the article lists the hydrocarbons separated from a Ponca, Okla., crude petroleum. The principles involved in the purification of hydrocarbons are briefly reviewed, and the processes of purification described. Table 2 lists the A.P.I.-N.B.S. Hydrocarbons completed July 1, 1943, to June 30, 1947. One hundred and twenty pure hydrocarbon compounds were separated and purified in this period. Criteria of purity are described. A bibliography of 19 references is included.

Vanadium Oxide, A Hydrogenation Catalyst. V. I. KOMAREWSKY, L. B. BOS AND J. R. COLEY, *Jour Am. Chem. Soc.*, **70** (1948) pp. 428-30.

Vanadium oxide has been widely used as an oxidation catalyst. It has also been reported to be an effective catalyst for the dehydrogenation of hydrocarbons. The work reporting a direct hydrogenation of unsaturated hydrocarbons in the presence of vanadium oxide catalyst is described. Vanadium oxide-aluminum oxide catalysts were prepared by the coprecipitation from solutions of sodium vanadate and aluminum nitrate. The per cent hydrogenation of isobutylene, hexene-1 and octene-1 obtained was over 95 percent. The optimum temperature for the olefin hydrogenation was found to be 400° C. The space velocity is critical for this reaction. Acetylene was hydrogenated successfully, the reaction product consisting of 92.9 percent ethane, 0.2 percent acetylene and 6.9 percent ethylene. Attemps to hydrogenate benzene were unsuccessful. However, at 115 atmospheres pressure 27 percent cyclohexane was obtained. An important feature of the vanadium oxide catalyst is its resisttance to sulfur poisoning.

Pyrolysis of Methane Flowing Through a Porcelain Tube in the Region 1000°. ALVIN S. GORDON, *Jour. Amer. Chem. Soc.*, **70** (1948) pp. 395-401.

The decomposition of methane was studied in a porcelain tube at temperatures between 1007 and 1075° C. Within this range of temperature the first-order reaction constants increased with increasing decomposition. There was a catalytic effect of acetylene, probably homogeneous, which accounts for the first-order reaction because the percentage of acetylene increased with increasing percentage of decomposition. The author concluded that ethylene had probably no effect on the decomposition of methane. The surface initially had a large catalytic effect but this disappeared by the time 30 percent of the methane was decomposed. No induction period was observed. In the early stages of the decomposition, 50 percent nitrogen or steam in the entering gas resulted in approximately the same percentage of decomposition of the methane for equal contact time.

The Preparation of Optically Active Hydrocarbons by the Wurtz Reaction. R. L. LETSINGER, *Jour. Amer. Chem. Soc.*, **70** (1948) pp. 406-9.

(—)1-Phenyl-2-methylbutane and (+) 1-phenyl-2-methyloctane were prepared in good yields by the reaction of benzylsodium with (+)2-bromobutane and

(+)2-bromoöctane, respectively. (+)1-Phenyl-2-methylbutane was prepared by the condensation of phenylsodium with (+)2-methyl-1-bromobutane. The extent of racemization in the reaction of benzylsodium and 2-bromobutane was estimated and the stereochemical relationships are discussed.

The Catalytic Decomposition of Ethane and Ethane-Hydrogen Mixtures. CHARLES KEMBALL AND HUGH STOTT TAYLOR, *Jour. Amer. Chem. Soc.*, **70** (1948) pp. 345-51.

The mass spectrometer can be used to follow the reaction of ethane and hydrogen to produce methane. The ethane-hydrogen decomposition on a supported nickel catalyst follows equation

$$\text{Rate} = k p_{C_2H_6}^{0.7} p_{H_2}^{-1.2}$$

over a wide range of composition. Energy of activation is 52 kcal./mole. At low ratios of hydrogen-ethane, abnormal kinetics, depending on a high inverse power of the hydrogen pressure, is observed. This is ascribed to heterogeneity of the surface. Ethane will decompose quantitatively over a nickel catalyst to methane and carbon. The rate is proportional to $p_{C_2H_6}^{0.7}$ and energy of activation is 40 kcal./mole. The carbon is reversibly deposited and can be converted to methane by hydrogen. The rate-determining step of methane formation is the breaking of the carbon-carbon bond. Ethane-hydrogen reaction and ethane decomposition can be interpreted by a mechanism involving the presence on the surface of adsorbed ethylene, in equilibrium with adsorbed ethyl radicals and hydrogen. The fact that the initial rate of the production of methane is greater from ethylene than from ethane substantiates this view.

Low-Temperature Polymerization of Isobutene by Titanium Chloride. P. H. PLESCH, *Nature*, **160** (1947) pp. 868-9.

The low-temperature polymerization of isobutene in hexane solution by titanium chloride ($TiCl_4$) requires a co-catalyst. The co-catalyst soluble in hexane and which would form a $TiCl_4$ complex also soluble in hexane was desired. Dry trichloracetic acid in hexane solution was found to be efficient for the purpose. For a starting temperature near −80° C. the maximum rate of polymerization under adiabatic conditions is a function of the concentrations of isobutene, $TiCl_4$ and trichloracetic acid.

Condensation of Saturated Hydrocarbons with Haloölefins. II. The Reaction of Isobutane with cis- and trans-Dichloroethylene. LOUIS SCHMERLING, *Jour. Amer. Chem. Soc.*, **70** (1948) pp. 379-81.

1,1,2-trichloro-3,3-dimethylbutane, is the principal product of the reaction of isobutane with 1,2-dichloroethylene in the presence of aluminum chloride at room temperature. It is obtained in 35 percent yield when cis-dichloroethylene is used, and in less than 5 percent yield when trans-dichloroethylene is used. The probable mechanism of the reaction is given. It is shown that the reaction serves as a means of evaluating recently proposed mechanisms for the alkylation of isoparaffins with olefins.

Perfluorinated Olefins. ALBERT L. HENNE AND THOS. H. NEWBY, *Jour. Amer. Chem. Soc.*, **70** (1948) pp. 130-32.

The synthesis of perfluorinated ethyl-

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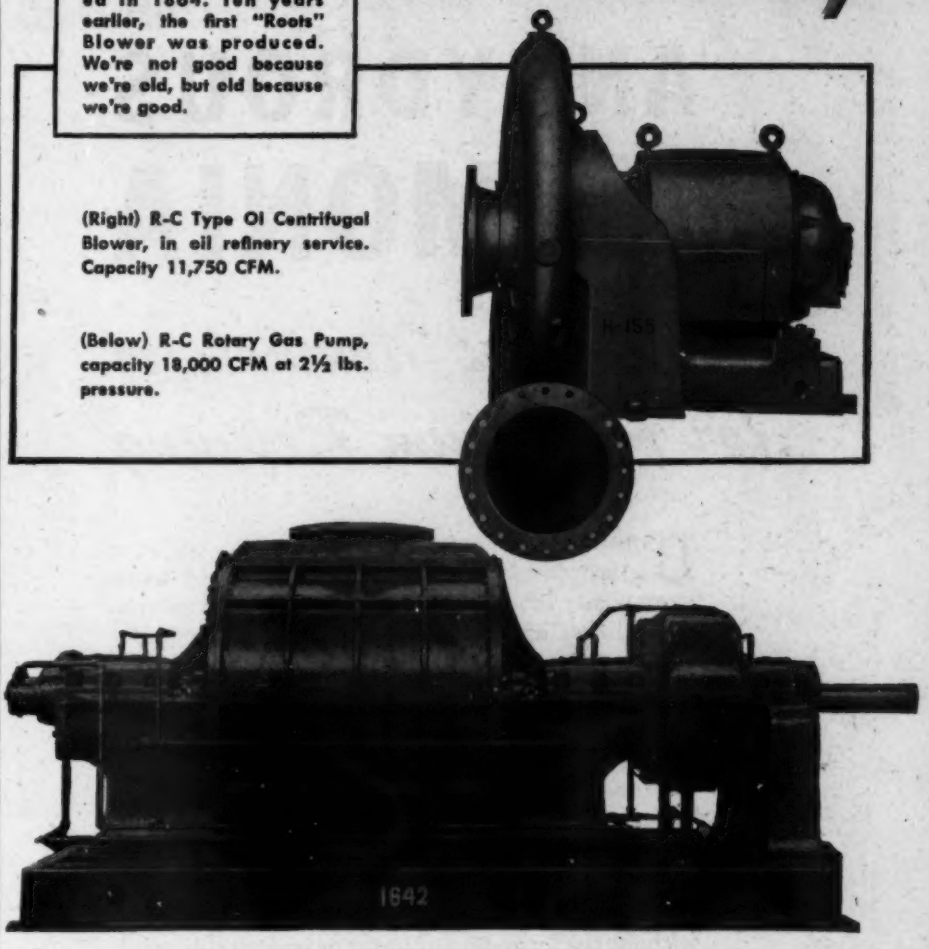

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ene and its polymerization to commercial TEFLON, $(CF_2)_n$, are now industrial processes. Efforts of the authors were directed towards the synthesis of other fully fluorinated olefins in order to study their polymerization characteristics and their ability to copolymerize with $CF_2{=}{=}CF_2$. A synthesis sequence for $CF_3CF{=}{=}CF_2$ was complicated by a rearrangement. A practical, simple sequence gave $CF_3CF{=}{=}CFCF_3$. The preparation of $C_2F_5CF{=}{=}CF_2$ was defeated by rearrangements.

Manufacture: Processes and Plant

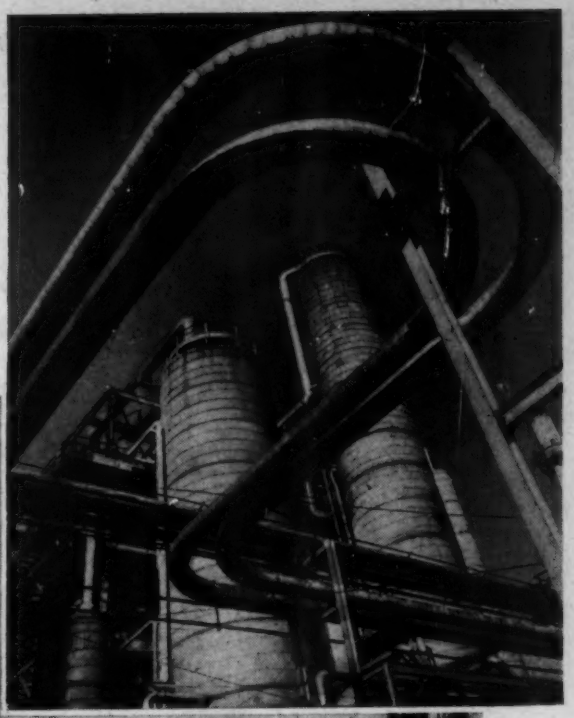
Pressure Drop Across Tube Banks. D. F. BOUCHER AND C. E. LAPPLE, *Chem. Eng. Progress*, **44** (1948) pp. 117-34.

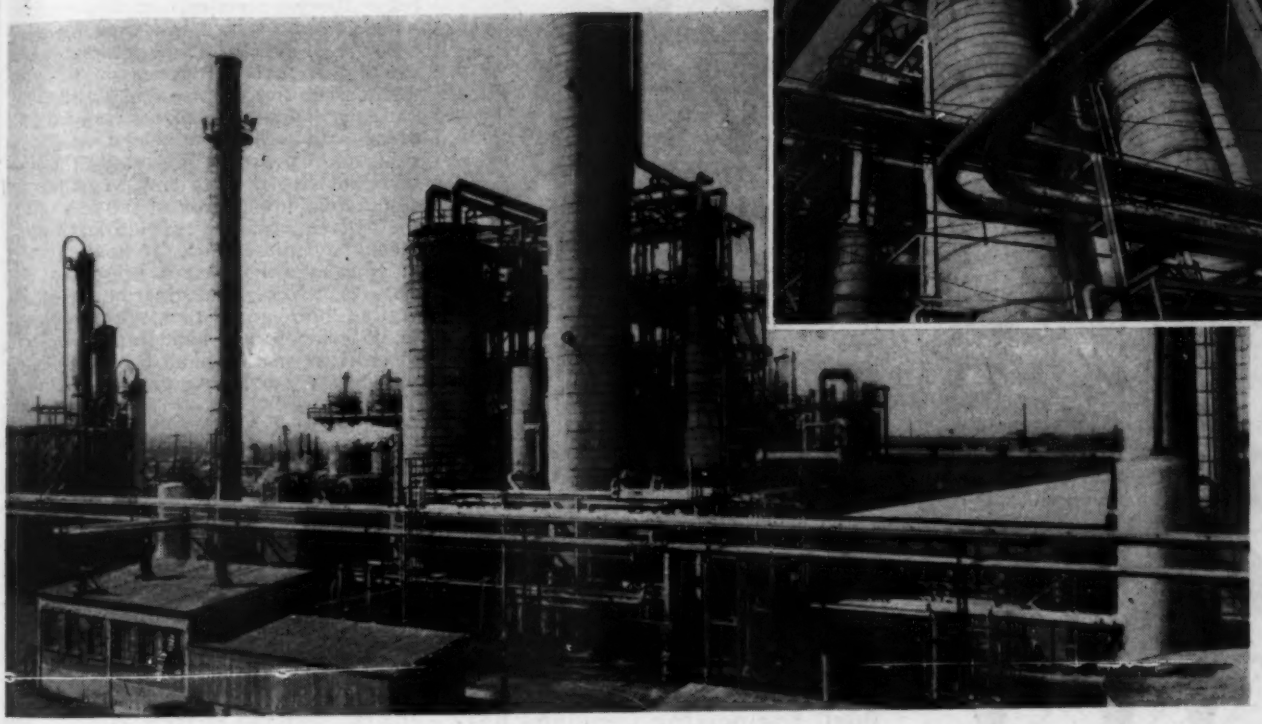
The estimation of pressure drop across tube banks is a field in which the data are conflicting and in which correlations are needed. The purpose of the paper is to compare the methods that have been proposed in the light of all the data now available. Only plain tubes are considered. Incorporation of allowances for tube-bank configuration in a single friction factor-Reynolds number relationship is fundamentally unsound and any attempt to use such a factor is purely empirical. A fundamental correlation should be based on isothermal and isobaric data. All availabel data for turbulent flow are summarized on a consistent basis in Tables 2 and 3, together with details of equipment and operation. An indication of the effect of the number of rows and tubes on the friction factor for turbulent flow is given in Table 5. The graphical correlation of Grimison, in which friction factor is plotted against Reynolds number with transverse and longitudinal spacings as parametric curves, appears to be the best available method for a general representation of data for both in-line and staggered arrangements over a Reynolds number range of 2,000 to 40,000 and involves an average deviation from the data of ± 15 percent. A bibliography of 37 references is included.

Reaction Rate, Heat Transfer, and Temperature Distribution in Fixed-Bed Catalytic Converters. Solution by Electrical Networks. R. H. WILHELM, W. C. JOHNSON, R. WYNKOOP AND D. W. COLLIER, *Chem. Eng. Progress*, **44** (1948) pp. 105-16.

The literature with regard to thermal problems in tubular pseudo-isothermal catalytic converters in which the temperature of the reactor wall is controlled is reviewed. Fundamental heat-balance equations are given. Data needed for solution of the equations are discussed under the headings of 1) kinetics and 2) solid-gas convection heat and mass-transfer coefficients, and 3) thermal conductivity of granular media. A method for rapidly analyzing thermal problems in catalytic reactors by means of an electrical network is described. The network depends upon the mathematical analogy between Fourier's law of heat conduction and Ohm's law of electrical conduction. No limitations are imposed in the use of the network with respect to the form of temperature in the use of the network with respect to the form of temperature dependencies or variation

TIDE WATER ASSOCIATED OIL CO. USES *Super Nickel 702* FOR RETUBING AT BAYONNE





General views of Tide Water Associated Oil Company Refinery at Bayonne, New Jersey.

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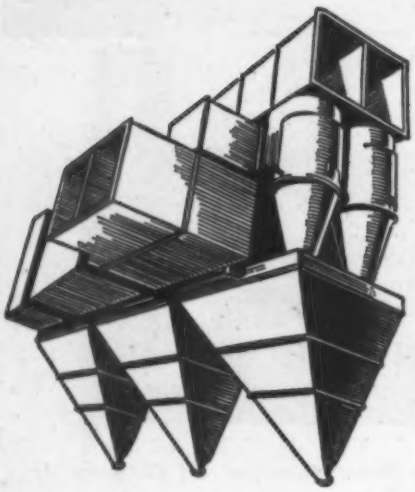

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of conditions from point to point. Construction of a given network is described. The paper provides a theoretical background for experimental studies of catalytic reactors in which the thermal problems are important. A bibliography of 25 references is included.

Furfural as a Selective Solvent in Petroleum Refining. L. C. KEMP, JR., G. B. HAMILTON AND H. H. GROSS, *Ind. Eng. Chem.*, **40** (1948) pp. 220-7.

Furfural has a high solvent power for those components of petroleum that are relatively unstable to oxygen, as well as other undesirable materials including color bodies, resins, carbon-forming bodies, and yet, at the working temperatures used, has a low solvent power for the desirable components. There are today some twenty plants using furfural, and ten additional plants are in the process of being engineered and constructed. The paper discusses the furfural refining process and describes the commercial application of the method to the solvent refining of lubricating oil stocks and diesel fuels. A flow sheet of a typical refining unit is presented. Results of the furfural treatments of the various lubricating oil stocks are given in tabular form, including paraffinic, mixed-base, and asphalt-base crudes. A bibliography of 14 references is included.

Commercial Application of Hydrocarbon Synthesis in the United States. JAMES E. LATTA AND SCOTT W. WALKER, *Chem. Eng. Progress*, **44** (1948) pp. 173-76.

Petroleum resources have been used in enormous quantities and while estimates of the ultimate supply vary widely, it is evident that crude petroleum is becoming scarcer and more expensive and will be exhausted within several more decades. These facts have given importance to consideration of other sources of petroleum-type hydrocarbons, and the synthesis of these hydrocarbons from hydrogen and carbon monoxide is one of several possibilities. Announcement has been made of plans for two commercial plants to synthesize hydrocarbons and their oxygenated derivatives. The synthesis gas is prepared by the partial combustion of natural gas with oxygen at 20 atmospheres pressure. The reaction is conducted over a cheap iron catalyst in a fluid bed yielding hydrocarbons as the principal by-products, with smaller amounts of oxygenated derivatives. The synthetic hydrocarbons are recovered, and, by relatively simple refining processes, are converted to high quality gasoline and heavier fuels. Oxygenated compounds are obtained in complex mixtures that are processed for the separation of individual components. Some of these materials are important raw materials for chemical syntheses and for use as solvents, while others have never been available previously in commercial quantities, and applications for them must be developed. Two commercial plants incorporate the results of five years of intensive research on the problems involved. Operability of the fluid iron catalyst process has been confirmed in pilot plants, but much work will be required to determine the many effects of changes in the reaction variables. Successful operation of these two large plants and the development of a more comprehensive understanding of the synthesis reaction mechanism through continued research will vitally affect the subsequent installation of additional

plants utilizing natural gas and, eventually, coal as raw materials for the synthesis of hydrocarbons and their derivatives.

Products: Properties, Utilization and Analysis

Insecticidal Efficiency of Saturated Petroleum Fractions. G. W. PEARCE, P. J. CHAPMAN AND D. E. H. FREAR, *Ind. Eng. Chem.*, **40** (1948) pp. 284-93.

At least two basic factors appear to be involved in the insecticidal properties of petroleum fractions employed for controlling insect and mite pests on fruit trees, namely, structural composition and molecular weight of the hydrocarbons present. The work reported in the paper was undertaken because the relationship between the molecular weight of petroleum hydrocarbons and their insecticidal efficiency has not been determined. Two series of narrow boiling petroleum fractions were prepared by fractional distillation of two technical white oils of commercial origin. The oils selected for fractionation represent distinctly paraffinic and naphthenic types, the viscosities of which fall within the spray oil range. The fractions obtained were tested for their insecticidal efficiency using the eggs of the oriental fruit moth, Grapholitha molesta Busck, as test material. Correlations between insecticidal efficiency and various properties of the fractions, such as molecular weight, viscosity, density, refractive index, aniline point, and structural character were observed. It is evident from these correlations that conventional spray oils may contain a high proportion of hydrocarbons that are inactive insecticidally. Use of the data as a basis for establishing the properties of an ideal composition as well as their use as standards for a relative evaluation of oils as insecticides is indicated. The properties of an ideal oil are Saybolt viscosity at 100° F. 70-80 seconds, unsulfonated residue not less than 92 percent, density 0.840 (max.), boiling point range at 1 mm. Hg. 10° ± 5° F., 50 percent boiling point at 1 mm. Hg. 370° ± 5° F., molecular weight 340 ± 5, pour point not greater than 30° F. A bibliography of 26 references is included.

Oil-Paper Dielectrics. Effect of Asphalt as a Contaminant in Oil and in Impregnated Paper. JOHN D. PIPER, PORTIA TREEND AND KATHRYN S. BEVIS, *Ind. Eng. Chem.*, **40** (1948) pp. 323-28.

Results of the investigations to date have been summarized as follows: Oil-soluble substances of wide variety, including most medium molecular weight acids, alcohols, aldehydes, esters, ketones, and peroxides, have little effect on power factors and conductivities of oil or oil-impregnated paper. All additives thus far found cause high power factors and conductivities in insulating oils or oil-impregnated papers and fall either in Class A or Class B additives. Class A additives contain at least one component that is incompletely soluble in oil. Class B additives belong to the type of substances, notably acids, that in water or other solvents dissociate electrolytically. Only the strongest of these substances, such as sulfonic acids and salts of organic bases, cause an increase in the power factor of the oil that is of practi-

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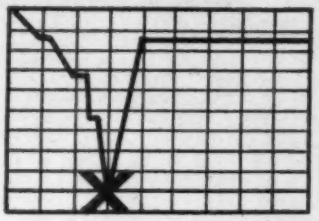

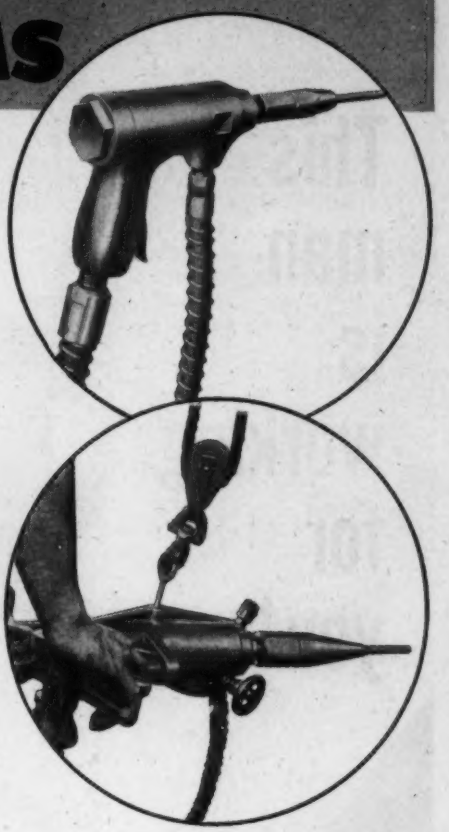

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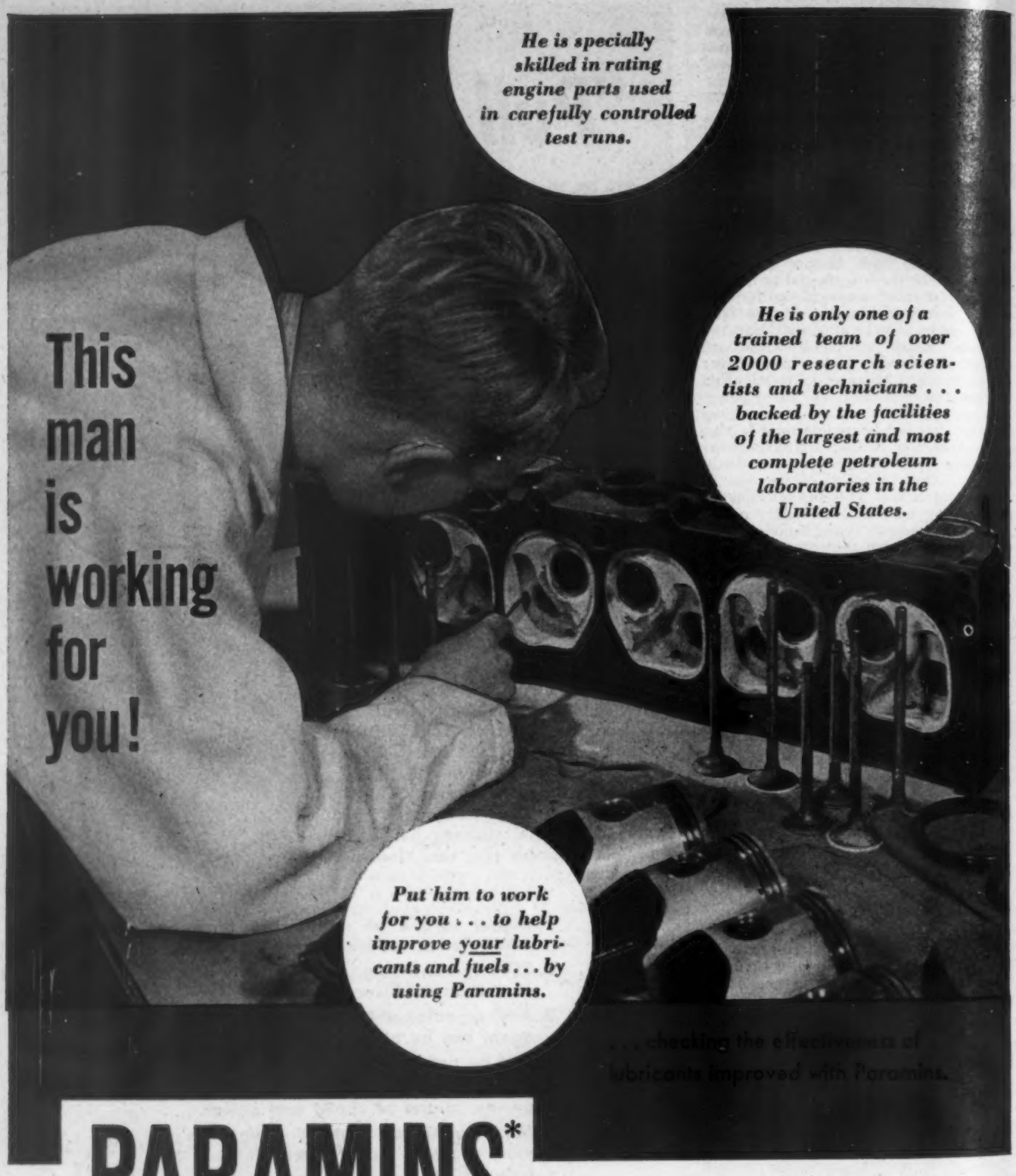
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cal significance. Addition of asphalt to an insulating oil increases the power factor and conductivity. This conductivity is nearly independent of frequency. The dielectric properties of the asphalt solutions are not markedly different from those of solutions of an electrically dissociable material, such as lauryl sulfonic acid. Although asphalt, like lauryl sulfonic acid, is strongly sorbed from oil by paper, the dielectric properties of paper impregnated with solutions of these contaminants are strikingly different. The authors suggest a means for determining whether high conductivity in an insulating oil is caused primarily by electrolytically dissociable materials, such as acids, or by colloidally dispersed materials, such as asphalt. A bibliography of 9 references is included.

Specific Heats of Volatile Liquids. G. C. WILLIAMS, *Ind. Eng. Chem.* **40** (1948) pp. 340-1.

A procedure is described for determining specific heats of a liquid of a relatively high volatility. When the specific heats are needed at temperatures close to the boiling point, the usual methods of specific heat determination are inaccurate. An equation is given by which it is possible to relate specific heats to the weight of the substance employed, the heat equivalent of the calorimeter, and three rate expressions: heat input, measured temperature rise, and temperature change at no-heat period. Data for a typical measurement are given and the calculation of a specific heat is illustrated.

Mass Spectrometer-Hydrochlorination Analysis of Butenes. F. W. MELPOLDER AND R. A. BROWN, *Anal. Chem.* **20** (1948) pp. 139-42.

To improve the accuracy of determining individual butenes, isobutene is converted to tert-butyl chloride prior to mass spectrometer analysis. Two special methods of computation were devised, with modifications of the basic principles of analyses proposed by Washburn, Wiley, and Rock. The procedure used in the work, the methods of computation and the data secured are presented in considerable detail.

Quantitative Determination of Dicyclopentadiene. FELIX BERGMANN AND HELENE JAPHE, *Anal. Chem.*, **20** (1948) pp. 146-7.

Although dicyclopentadiene and its monomer have become of considerable industrial importance in recent years, the few methods that have been developed for the quantitative determination of the dimeric hydrocarbon are subject to limitations. The authors describe a method for the quantitative analysis of dicyclopentadiene which involves the addition of formic acid to one of the double bonds and determination of the saponification equivalent of the resulting formate ester. The method is also applicable to the analysis of the higher polymers of the hydrocarbon. The analytical procedure is described in considerable detail and the results of typical analyses are given.

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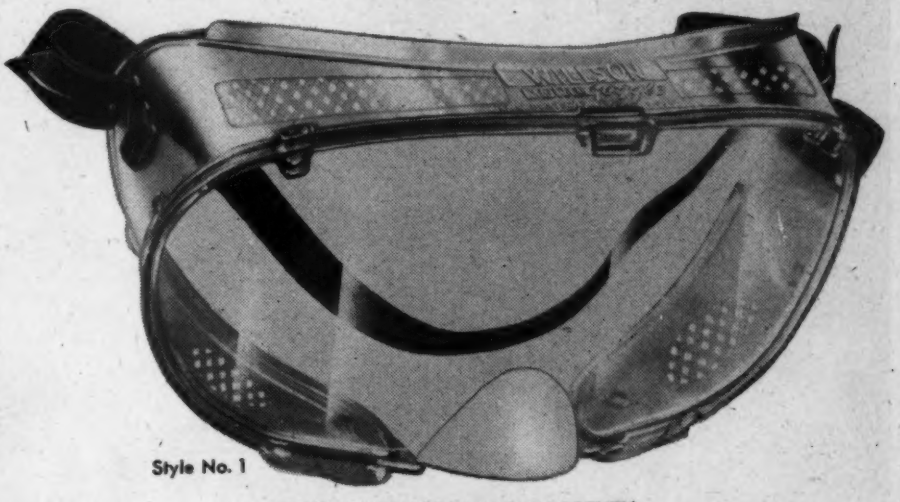

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A terrific oil tank explosion in the Signal Hill oil field early today caused damage estimated at $50,000 and threatened the lives of firemen who battled the flames which followed.

One tank containing 5000 gallons of oil was destroyed, a second was damaged and a nearby derrick was wrecked in the blast.

It was believed an accumulation of fumes above the crude oil level was responsible for fire in the first tank.

Risking their lives, a group of firemen entered the second tank, also containing 5000 gallons of oil, and extinguished the flames with foam and water. There were no injuries.

The tank exploded with a roar that could be heard for miles around. Flames shot 150 feet into the air.

Three ignited wood derricks were saved from destruction.

A high voltage line was destroyed.

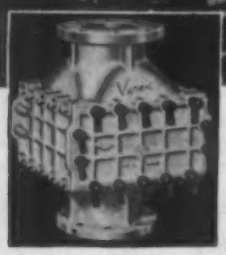
Two police officers saw the tank explode as they were giving a traffic citation nearby. They turned in an alarm that enabled firemen to reach the scene before the blaze had spread too much.

The Signal Hill fire department was assisted by three companies of the Long Beach fire department.

Flames into th...

P2

Photo by L.A. Daily News



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AND COVERING PATENTS PUBLISHED IN THE OFFICIAL GAZETTE DATED JANUARY 13, 20, 27 AND FEBRUARY 3, 1948

REFINING

U.S.P. Re 22,963. Process for Resolving Oil-in-Water Emulsions. L. T. Monson, W. A. Anderson, and F. W. Jenkins to Petrolite Corporation, Ltd.

Petroleum emulsions of the oil-in-water type are broken by subjecting them to the action of a surface-active heat-polymerized amino-alcohol of the formula $[HOR.(OR)_m]n - N - [R_1]n'$, wherein OR is an alykylene oxide radical with not over 4C atoms selected from the radicals of ethylene oxide, propylene oxide, butylene oxide, glycide, and methyl glycide, R_1 is hydrogen or an alkyl radical with 6 or less C atoms, m is a numeral from 1-3, n is 2 or 3, n' is 0 or 1 and $n + n' = 3$. The heat-polymerized compound is either the anhydro base, or the hydrated base, or a salt. About 1 percent of the demulsifier is preferably used. Heat-polymers of triethanolamine can, for example, be employed.

U.S.P. 2,434,322. Azeotropic Distillation of Toluene with Propionic Acid. J. W. Latchum, Jr., and J. S. Connors to Phillips Petroleum Company.

Toluene is separated from a hydrocarbon fraction boiling within a boiling range of up to about 30° F. (e. g. of the range of 210-240° F.) and containing toluene and non-aromatic hydrocarbons by continuously distilling this fraction in the presence of propionic acid. Toluene and a portion of the propionic acid are thereby concentrated in the still bottoms. The propionic acid is water extracted from these bottoms as well as from the non-aromatic overhead hydrocarbons, dehydrated and recycled.

U.S.P. 2,434,528. Method for Removing Ash-Forming Constituents and Sediment from Alkaline Petroleum Residuum. E. F. Wadley to Standard Oil Development Company.

Alkaline petroleum residua obtained by distilling hydrocarbon oil in contact with alkali are mixed with 50-150 percent by vol. of water H_2SO_4 of a strength of about 3N-10N is added in an amount sufficient only to neutralize the residua. A petroleum sulfonate is added in an amount of 0.05-0.3 percent by weight of the water, the materials are mixed thoroughly at elevated temperature (e.g. 125-212° F.) and then allowed to settle and stratify. The oil layer formed is withdrawn. It is substantially free of ashes. The petroleum sulfonate acts as a demulsifier.

U.S.P. 2,434,578. Antiknock Motor Fuel. P. Miller to Standard Oil Development Company.

The antiknock motor fuel according to this patent consists of a gasoline fuel containing a small amount of a stable complex of the formula $M'X_x.M''Y_y.(N')_n$ wherein M' and M'' represent different central heavy metal constituents of groups I, IV, VI, VII, or VIII of the periodic table, X and Y are halogen atoms taken x and y times to satisfy ordinary valences of M' and M'', respectively, and $(N')_n$ represents molecules of CO, NO, CN, nitrogen bases, or an oxide of said metals having coordinate linkages to the metal constituents, n being the number of these molecules corresponding to an available coordination valence of said metals. An example of such a complex is: $Fe(CO)_4Cl_2SnCl_2$. It has an octane number increasing effect.

U.S.P. 2,434,402. Method of Concentrating Dilute Aqueous Solutions of Metal Ammonium Complex Compounds. A. W. Fleer to Shell Development Company.

The patent relates to the concentration of dilute aqueous solutions of ammonium complexes of metal salts, particularly of aqueous copper ammonium acetate solutions as obtained in the separation of butadiene from mixtures with other hydrocarbons. The dilute solution is heated below its boiling point to remove mixed water-NH_3 vapor while adding supplemental ammonia with a NH_3 content greater than that of the mixed vapor, the amount of introduced NH_3 being about equal to that removed with the vapor. The deposition of insoluble products during the concentration is thus avoided.

U.S.P. 2,434,409. Process for Purifying a Hydrocarbon Mixture Containing Small Amounts of Organic Fluorine Compounds. V. N. Ipatieff and C. B. Linn to Universal Oil Products Company.

Hydrocarbon mixtures are freed from small amounts of organic fluorine compounds contained therein and deriving from the alkylation of an isoparaffin and an olefin in the presence of HF or the like, by treating the mixture with H_2SO_4 of at least 90 percent concentration.

U.S.P. 2,434,424. Plural Stage Extractive Distillation with Inverse Solvents. R. C. Morris and T. W. Evans to Shell Development Company.

Butadiene can be separated from a C_4 hydrocarbon mixture also containing beta butylene and more saturated hydrocarbons by extractively distilling this mixture in the presence of a first relatively high boiling polar solvent to produce a first overhead fraction comprising the more saturated hydrocarbons and a residual fraction comprising butadiene, beta butylene and the solvent. The latter is separated and at least a part of the remaining residual fraction is then extractively distilled in the presence of a relatively high boiling non-polar solvent for beta butylene under conditions to take butadiene overhead. This method of separating unsaturated from more saturated hydrocarbons of a vaporizable mixture by successive use of a polar and a non-polar solvent is also applicable to the treatment of other hydrocarbon mixtures, e.g. to that of a mixture of aromatics and more saturated hydrocarbons.

U.S.P. 2,434,650. Motor Fuels and Preparation Thereof. W. A. Herbst to Standard Oil Development Company.

The motor fuel according to this patent comprises gasoline hydrocarbons, a knock reducing amount of aromatic amine, a gum inhibiting amount of alkylated hydroxy aromatic oxidation inhibitor free from any amino substituents, and a sufficient amount of CS_2 to stabilize the aromatic amine present in the fuel against discoloration during storage. The proportions can, for example, be: at least 50 percent by vol. of hydrocarbon stock, 0.5-0.6 percent of antiknock agent, 0.1-1.0 lb./1000 gallons of gum inhibitor, and less than 0.1 percent of CS_2.

U.S.P. 2,434,796. Process for the Extractive Distillation of C_4 Hydrocarbons with Furfural. K. H. Hachmuth to Phillips Petroleum Company.

The patent is concerned with a process for the separation of low-boiling aliphatic C_4 hydrocarbons of various degrees of saturation. The more unsaturated portion, e.g. butadiene, is separated from the more saturated portion. Details are given of a process for effecting this separation by fractional distillation in the presence of furfural as a selective solvent for the more unsaturated hydrocarbons.

U.S.P. 2,434,839. Process for Treating Hydrocarbons. O. L. Davis and A. C. Nixon to Shell Development Company.

In the treatment of a hydrocarbon mixture or fraction with an aqueous copper ammonium acetate solution, in which two phases are produced, a relatively stable dispersion of one phase in the other is obtained by the action of impurities formed as a result of the reaction of unsaturates in said mixture or fraction with each other and with oxygen. To prevent the formation of such a dispersion a relatively small (e.g. 0.0001-1 percent by vol. of the mixture) of an unsubstituted aliphatic alcohol with 8-14 C atoms per mol., miscible or soluble or spontaneously dispersable colloidally in the aqueous copper ammonium acetate solution is added to the reactants. This method can also be employed where other alkaline alkali metal or alkaline ammonia compounds capable of forming two phases with hydrocarbon mixtures are used. Decyl alcohol, lauryl alcohol, and 2-ethyl hexanol are examples of the alcohols useful in this process.

• MORE PATENTS PAGE 221

CATALYST PREPARATION, ACTIVATION, AND REGENERATION

U.S.P. 2,434,418. Preparation of Magnesium Silicate Adsorbents. W. A. LaLande, Jr., to Attapulgus Clay Company.

An alkali metal silicate and a water soluble magnesium salt are reacted in aqueous solution in the presence of a compound yielding NH_4 ions in aqueous solution, such as NH_4Cl, at a temperature of 150-400° F. The insoluble magnesium silicate formed is separated. The product obtained has superior decolorizing and catalytic properties.

U.S.P. 2,434,419. Process of Regenerating Adsorbents. C. D. Laughlin and H. M. Gwyn, Jr., to Attapulgus Clay Company.

A solid adsorbent, such as bauxite, containing moisture adsorbed from a wet combustible gas is regenerated by passing the wet gas through the wetted adsorbent and burning at least a portion of this wet gas. The resulting combustion products are passed in indirect heat exchange relationship with the adsorbent to increase its temperature sufficiently to drive off adsorbed moisture during the passage of wet gas therethrough.

U.S.P. 2,434,567. Method and Apparatus for Contacting Hydrocarbons with Catalyst Particles. C. E. Jahnig and C. W. Tyson to Standard Oil Development Company.

A reactant (e.g. hydrocarbons) is contacted with hot powdered catalyst in a reaction zone. Spent catalyst withdrawn from the bottom of this zone is passed to the bottom of a first regeneration zone for partial regeneration and then to a second regeneration zone arranged above the reaction zone for further regeneration. A fluidized column of catalyst is provided and a hydrostatic pressure is produced between the second regeneration zone and the reaction zone for passing regenerated catalyst to the reaction zone.

U.S.P. 2,435,379. Preparation of Spheroidal Catalyst. R. C. Archibald to Shell Development Company.

To prepare substantially spheroidal catalyst particles suitable for use in moving-bed catalyst reaction systems, an acidic, inorganic sol capable of forming a gel in the form of dispersed droplets is contacted with a body of liquid immiscible with the sol and containing a coagulating agent, such as ammonia. The droplets formed are maintained dispersed in the liquid for a time sufficient to convert the sol droplets into relatively stable gel particles.

U.S.P. 2,435,380. Process for Preparing Catalysts. R. C. Archibald and R. A. Trimble to Shell Development Company.

A catalyst of mixed sulfides of an element of the left column of group VI of the periodic table and of a metal of the iron group is prepared by treating an aqueous alkaline solution of a salt of a thioacid of that element with an aqueous solution of a soluble salt of that metal and with water-soluble acid to precipitate the sulfides of said element and of said metal. The thioacid solution must

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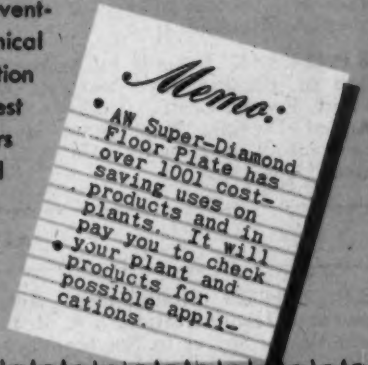
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contain sulfide ions in an amount at least equivalent to said metal and shall be employed in sufficient amount to obtain a pH of 1-2. Precipitation is carried out at a temperature below 50° C. The mixed sulfides obtained can be dried and subjected to reducing conditions. Mixed tungsten-nickel sulfides can, for example, be obtained. A catalyst specially suitable for the dehydrogenation of hydrocarbons of the type of cyclohexane naphthenes is obtained by the reduction of such a mixed sulfide until it has the composition: W·1.5-2 Ni·3-4S.

U.S.P. 2,434,602. Regeneration of Solid Materials. W. I. Thompson to Standard Oil Development Company.

Contact material containing combustible contaminants is suspended in a regenerating gas. The resulting suspension is passed successively in concurrent flow through a plurality of regenerating zones and the temperature of the catalyst is reduced between these zones. The contact material is also separated between these zones from the regenerating gas and then resuspended in a fresh stream of regenerating gas for further regeneration.

CRACKING AND REFORMING

U.S.P. 2,434,434. Hydrocarbon Conversion. D. Read, Jr., to Universal Oil Products Company.

A hydrocarbon oil is successively treated in two fluidized catalyst cracking stages with intermediate fractionation of the effluent of the first stage and return of the high boiling fraction containing entrained catalyst particles to this first stage. The fraction of gasoline boiling range from the effluent of the first stage forms the charge for the second catalytic treatment. The products of the second treatment are fractionated correspondingly and the high boiling fraction thereof containing entrained catalyst is returned to the second stage of the process.

U.S.P. 2,434,634. Production of High Octane Gasolines. J. R. Bates to Sun Oil Company.

A gasoline obtained by cracking a petroleum oil is treated to isomerize 1-olefins to 2-olefins and is then subjected to a second cracking treatment. If desired, the 1-olefins, e.g., 1-pentene, can be separated from the gasoline after the first cracking and then be isomerized to the corresponding 2-olefin which is recombined with the gasoline before the second cracking. A product of higher octane value is obtained by the intermediate isomerization of the 1-olefins.

U.S.P. 2,435,196. Cracking of Hydrocarbons with a Beryllium Phosphate Catalyst. G. R. Bond, Jr., and G. A. Mills to Houdry Process Corporation.

A hydrocarbon material, particularly an oil heavier than gasoline, is cracked in the presence of beryllium phosphate. This catalyst can be produced by precipitation of a soluble Be salt, e.g., Be sulfate, with an ammonium phosphate, forming the reaction product into pieces, and drying and calcining these pieces. Very high ratios of gasoline/gas and gasoline/coke are obtained by the use of the new catalyst under normal cracking conditions.

ISOMERIZATION

U.S.P. 2,434,338. Pariffin Hydrocarbon Isomerization. E. Solomon to The M. W. Kellogg Company.

The isomerization process according to this patent makes use of a normally solid volatile metallic halide catalyst which is vaporized and then mixed with the hydrocarbon vapors to be treated. Details of a process of this type are described and claimed.

U.S.P. 2,434,437. Process for Isomerization of Paraffin Hydrocarbons with Fluid Friedel-Crafts Catalyst and Added Dicyclic Naphthenic Hydrocarbons. W. E. Ross to Shell Development Company.

Pentane is converted to isopentane by contacting pentane in admixture with a promoting amount of a hydrogen halide under isomerizing conditions with a liquid mixture of decalin and a fluid $AlCl_3$-containing catalyst. Other saturated dicyclic alicyclic hydrocarbons can be used instead of the decalin and other Friedel-Crafts metal halides can be employed as the fluid catalyst. The method is also useful in the isomerization of other normal and branched chain paraffins with 4-10 C atoms. No substantial hydrocarbon decomposition takes place in these conversion operations.

ALKYLATION

U.S.P. 2,434,623. Simultaneous Alkylation and Desulfurization. J. R. Meadow and W. A. Stover to Socony-Vacuum Oil Company, Inc.

A mixture of gas oil which contains alkylatable constituents and has a relatively high S content and naphtha containing alkylatable constituents is treated with sufficient HF to ensure liquid phase separation of HF. A liquid HF portion is separated from a liquid desulfurized alkylated hydrocarbon portion and the latter is cracked. A low sulfur gasoline with increased gasoline to coke ratio is obtained.

U.S.P. 2,435,402. Production of Motor Fuels. S. H. McAllister and E. F. Bullard to Shell Development Company.

An excess of isoparaffin or alkyl cycloparaffin is contacted with an olefin and conc. H_2SO_4 under alkylating conditions producing a saturated hydrocarbon. Benzene sulfonic acid can be added to the H_2SO_4 catalyst as a promoter.

U.S.P. 2,435,028. Alkylation Process. W. E. Bradley to Union Oil Company of California.

An isoparaffin, such as isobutane, and an olefin, such as a mixture of butylenes, are constinuously fed to an alkylation system and contacted therein with a catalyst, such as H_2SO_4, under alkylation conditions. On settling the reaction products, a catalyst phase is separated from an alkylate phase containing a high proportion of isoparaffin. This alkylate phase is used for prediluting the olefin feed prior to contacting it with the catalyst.

U.S.P. 2,435,029. Process for Alkylation of Isoparaffins with Olefins. W. E. Bradley to Union Oil Company of California.

The patent describes a continuous method for reacting isoparaffins or aromatics with monoolefins in the presence of an alkylation catalyst and for producing gasoline of high octane number by such alkylation. The reagent mixture is agitated during the contacting operation with the catalyst. The molecular ratio of alkylatable hydrocarbon to olefin is preferably maintained at at least about 50:1 at the point where the olefin initially contacts the catalyst.

U.S.P. 2,435,038. Catalytic Dealkylation of Alkyl Aromatic Compounds. W. I. Gilbert and C. W. Montgomery to Gulf Research & Development Company.

A mono-hydric tertiary-butyl phenol such as mono- or di-tertiary-butyl cresol is contacted with an active clay catalyst at a temperature of 250-350° C. Isobutylene is obtained. The process is, for example, useful for the separation of meta- and para-cresol by first alkylating their mixture with isobutylene, separating the resulting alkylation products from one another by fractionation and then debutylating one or both of them separately by the patented process.

POLYMERIZATION AND CONVERSION

U.S.P. 2,434,536. Emulsion Polymerization Process. E. Arundale to Standard Oil Development Company.

In the polymerization of a conjugated diolefin, e.g., in the interpolymerization of butadiene-1,3 and styrene, in aqueous emulsion, an amount of up to 50 percent of an aliphatic mercaptan modifier or a mixture of such modifiers is added which contains 8-16 C atoms per molecule. This addition is made stepwise during polymerization. A rubber-like material is obtained.

U.S.P. 2,434,552. Polymerization Process. N. M. Elmore and H. C. Evans to Standard Oil Development Company.

Isobutylene is polymerized at low temperature in the presence of 150-400 parts of a liquefied normally gaseous paraffinic hydrocarbon internal refrigerant per 100 parts of isobutylene. BF_3 dissolved in an alkyl chloride with 1-2 C atoms per molecule to a concentration of 10-50 percent is added to a portion of the refrigerant which contains 2-3 C atoms per mol., to produce a solution which contains 0.04-2.4 percent of BF_3 and 0.1-4 percent of BF_3 calculated on the amount of isobutylene to be polymerized. The isobutylene is mixed with the rest of the refrigerant and this mixture is added to the before mentioned BF_3 mixture for effecting polymerization. Oily liquid polymers with a mol. wt. above 1000 as well as rubber-like solids with a mol. wt. above 25,000 and products of medium mol. wt. and properties can be obtained.

U.S.P. 2,434,833. Process for Condensa-

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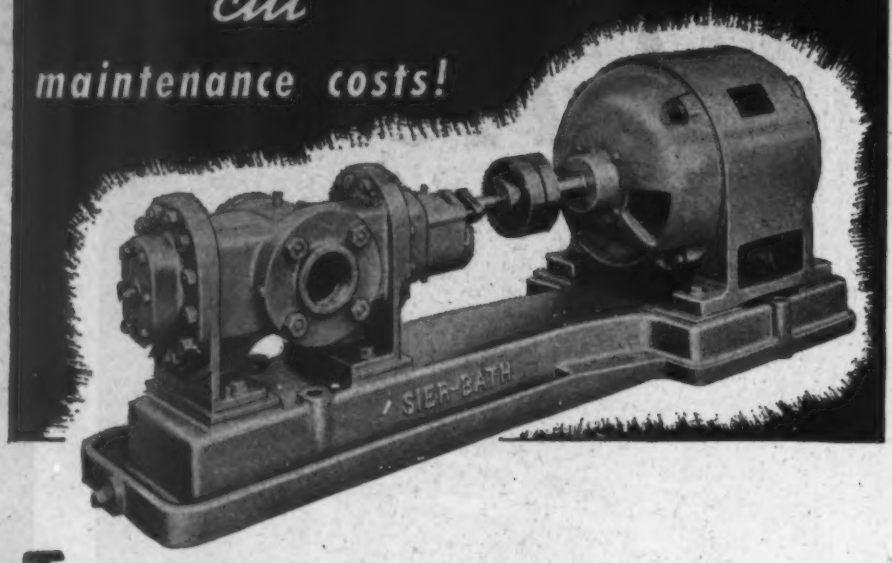

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tion of Hydrocarbons by Contact with Concentrated Sulfuric Acid in Silica Gel. F. G. Ciapetta to The Atlantic Refining Company.

In the conversion of hydrocarbons in the presence of a H_2SO_4 catalyst, e.g., in the polymerization of unsaturated hydrocarbons, such as isobutylene, or in the alkylation of an isoparaffin or an aromatic hydrocarbon with an olefin, the reactants are brought into intimate contact with silica gel having an initial water content of 3-15 percent by weight and which has been impregnated with SO_3 in a quantity sufficient to unite with the water content of the gel to form 100 percent H_2SO_4 in the pores of the gel. The resulting hydrocarbon polymers, alkylated isoparaffins, alkylated aromatic hydrocarbons, or the like are recovered.

U.S.P. 2,434,843. Method for Hydrocarbon Conversion in the Presence of Particle Form Catalysts. F. C. Fahnestock and W. Ulrich to Socony-Vacuum Oil Company, Inc., and The Lummus Company.

In the conversion of hydrocarbons in the presence a particle form catalyst which is passed as a substantially compact column cyclically through hydrocarbon conversion and a regeneration zone, fines resulting from the attrition of the catalyst material are removed by passing the stream of catalyst downwardly along an inclined, elongated chute so as to cause a concentration of the fines along the underside of this stream, and by withdrawing catalyst material containing some of the largest sized particles along with the concentration of the fines from the bottom of the chute near its lower end at a rate of only a minor fraction of the total rate of catalyst flow through the chute. After separation of the fines from the withdrawn catalyst material, the remaining coarse particles including said largest sized particles are returned to the circulating catalyst stream.

U.S.P. 2,435,158. Hydrocarbon Conversion Process. D. Read, Jr., to Universal Oil Products Company.

A stream of hydrocarbons which contains suspended spherical particles of catalyst is passed under conversion conditions through a reaction zone. The conversion products are fractionated after separation of the catalyst to separate a stream of normally gaseous products. The separated catalyst is suspended in at least a portion of these gaseous products and the resulting suspension is supplied to a catalyst regenerating zone where the catalyst is separated from the gaseous products and retained. The flow of the suspension to this zone is stopped after a predetermined time and the retained catalyst is regenerated at relatively low pressure. Another stream of gaseous products is then supplied under elevated pressure to the regenerating zone to force the catalyst into the feed stream of hydrocarbons to be converted. The process can, for example, be employed in cracking, reforming, and dehydrogenating operations.

U.S.P. 2,435,229. Kneader Polymerization Process for Olefins. M. D. Mann, Jr., to Jasco, Inc.

Isobutylene is mixed with an inert liquefied normally gaseous hydrocarbon refrigerant, e.g., dimethyl butadiene, to bring it to a temperature of —10 to —165° C. The mixture is subjected to

successive kneading treatments and polymerized in the presence of a Friedel-Crafts catalyst in fluid condition, such as BF_3.

U.S.P. 2,435,411. Diolefin Polymer Resins. F. J. Soday to The United Gas Improvement Company.

According to this patent a resinous product is obtained by polymerization in the presence of an acid-acting metallic halide of a light oil fraction free of more than 5 percent cyclopentadiene by weight and containing 50-80 percent diolefins, 20-48 percent amylenes, 2-5 percent pentanes, and 0.1-10 percent of a stabilizer obtained by reacting acetone with aniline, p-amino diphenyl, or diphenyl-p-phenylendiamine. The resulting resin is free from objectionable odor.

HYDROGENATION, DEHYDROGENATION, AROMATIZATION

U.S.P. 2,434,395. Preparation of Pure Aromatics from Petroleum Distillates. W. H. Claussen and T. M. Powell to California Research Corporation.

Pure aromatics, particularly pure toluene, are produced from naphthenic petroleum stocks boiling within the naphtha boiling range (e.g., at 180-235° F.) by treatment in the presence of dehydrogenation catalysts at increased temperature and pressure in the presence of a carrier gas containing free hydrogen. The resulting product is fractionally distilled to obtain a narrow cut aromatic fraction which is subjected to a second dehydrogenating catalytic treatment.

U.S.P. 2,434,522. Production of Diolefins. W. D. Seyfried to Standard Oil Development Company.

An olefin-containing hydrocarbon feed stock is passed through a dehydrogenation zone maintained under conditions to convert olefin into diolefin. The effluent is passed to a quenching zone where it is cooled to a temperature substantially above atmospheric by bringing it into intimate contact with a mineral oil containing free oxygen and an oxidation inhibitor. A phenol of petroleum origin, e.g., isobutyl-p-amino phenol can be used as the inhibitor and a naphthenic gas oil as the mineral oil. The diolefins formed, such as butadiene, are recovered. The presence of the oxidation inhibitor prevents the formation of polymers which are otherwise formed from the unsaturated hydrocarbons under the influence of peroxides formed by the action of the free oxygen present in the oil on the unsaturated compounds.

U.S.P. 2,434,537. Two-Stage Synthesis of Hydrocarbons. F. T. Barr and W. G. Scharmann to Standard Oil Development Company.

A mixture of CO and H_2 is forced into a first reaction zone containing a body of powdered catalyst in the form of a dense suspension. Contact time and temperature in this zone are controlled (e.g., to 350-500° F. at a pressure of 10-250 psi.) to induce substantial formation of hydrocarbons which are liquid at operating temperatures, limiting production to not more than about 60 percent of the amount attainable under the reaction conditions. The reaction products withdrawn from the first zone are discharged into a zone which contains at least one bed of stationary catalyst. The reaction is permitted to proceed in this second zone at synthesis temperatures until further substantial or maximum quantities of normally liquid products are formed.

U.S.P. 2,434,577. High-Compression Motor Fuels and Their Manufacture. P. Miller to Standard Oil Development Company.

A cracked gasoline of high olefin content and containing substantial amounts of aromatic and naphthenic hydrocarbons is fractionated to form a fraction boiling below 158° F. and a fraction boiling above 158° F. The lower boiling fraction is treated to replace paraffinic components thereof with mono-olefins with 4-6 C atoms until the mono-olefins constitute at least 75 percent by volume of this fraction and 10-30 percent by volume of the fuel. The fraction of the gasoline boiling above 158° F. is hydrogenated in the presence of a catalyst to reduce the olefin content to less than 10 percent by volume while maintaining the distillation characteristics substantially unaltered. The olefin-enriched fraction boiling below 158° F. is then recombined with the higher boiling hydrogenated fraction. A motor fuel with improved lead susceptibility and road octane rating is obtained.

U.S.P. 2,434,923. Distillation of Butadiene from Methyl Acetylene and Higher-Boiling Hydrocarbons. K. H. Hachmuth to Phillips Petroleum Company.

In the manufacture of butadiene by catalytic dehydrogenation of n-butenes hydrocarbon mixtures are obtained which in addition to butadiene contain methyl acetylene and at least one hydrocarbon less volatile than butadiene, e.g., butene-2. Butadiene is separated from such mixtures by treatment in a fractionator. An overhead vapor comprising butadiene and methyl acetylene and a bottom product comprising the less volatile hydrocarbon are obtained. The overhead vapor is condensed and a vaporous fraction containing methyl acetylene in considerably higher concentration than the other fractions is withdrawn from the condensate. The residual liquid condensate essentially consisting of butadiene and methyl acetylene is returned to the top of the fractionator as reflux and liquid butadiene is withdrawn from the fractionator at a point intermediate the points at which the feed and reflux enter. The resulting butadiene contains not over 0.05 percent by weight of methyl acetylene.

U.S.P. 2,435,443. Separation of Gem Cyclic Hydrocarbons from Nongem Cyclic Hydrocarbons by Selective Dehydrogenation. V. N. Ipatieff and H. Pines to Universal Oil Products Company.

A gem alkyl cyclohexane, i.e. a compound of this class which has at least one C atom in the ring with two substituent groups attached thereto, is separated from a non-gem alkyl cyclohexane by subjecting the mixture to dehydrogenation in the presence of a suitable catalyst, e.g. a platinum catalyst, at a temperature at which substantially only the non-gem compound is dehydrogenated to form an aromatic hydrocarbon and the gem compound is left substantially unchanged, preferably at a temperature below 275° C. The gem compound 1,1,-2,3-tetramethylcyclohexane can, for example, be separated by this method from the non-gem compound 1-methyl-4-isoproplycyclohexane.

U.S.P. 2,435,551. Production of Hydrocarbons from Carbon Monoxide and Hydrogen. J. F. Black to Standard Oil Development Company.

CO and H_2 are reacted at a temperature of 300-450° F. in the presence of a fluidized catalyst comprising nickel chromite prepared by combining at elevated temperature under agitation an aqueous solution of a nickel salt with an ammoniacal solution of ammonium chromate and separating, drying and calcining the precipate formed. The use of the novel catalyst makes unnecessary the otherwise required accurate temperature control of the Fischer-Tropsch process.

HEAVY OILS AND WAXES

U.S.P. 2,434,490. Rust Preventing Lubricating Oil Compositions. G. W. Duncan to Standard Oil Development Company.

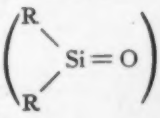
The rust preventing composition consists of a lubricating mineral oil and 0.5-10 percent by weight of a fatty acid monoester of sorbitan. The fatty acid radical of this ester shall have 8-30 C atoms. Sorbitan mono-laureate and mono-oleate are examples of the additive.

U.S.P. 2,434,539. Lubricants. A. Beerbower and A. J. Morway to Standard Oil Development Company.

The grease-like lubricant prepared according to this patent consists of a slurry of dehydrated and activated LiOH, $Ba(OH)_2$, or $Sr(OH)_2$ in mineral lubricating oil. The dehydration and activation of the slurry is effected by heating the hydroxide in the lubricating oil to a temperature above 225° F. A fatty acid, such as oleic acid or naphthenic acid, can be added in small amounts to form a soap with the metal hydroxide present.

U.S.P. 2,435,124. Method of Dispersing a Foam-Reducing Silicon Containing Compound. L. C. Bollinger to Shell Development Company.

A compound of the formula



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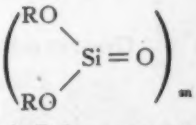
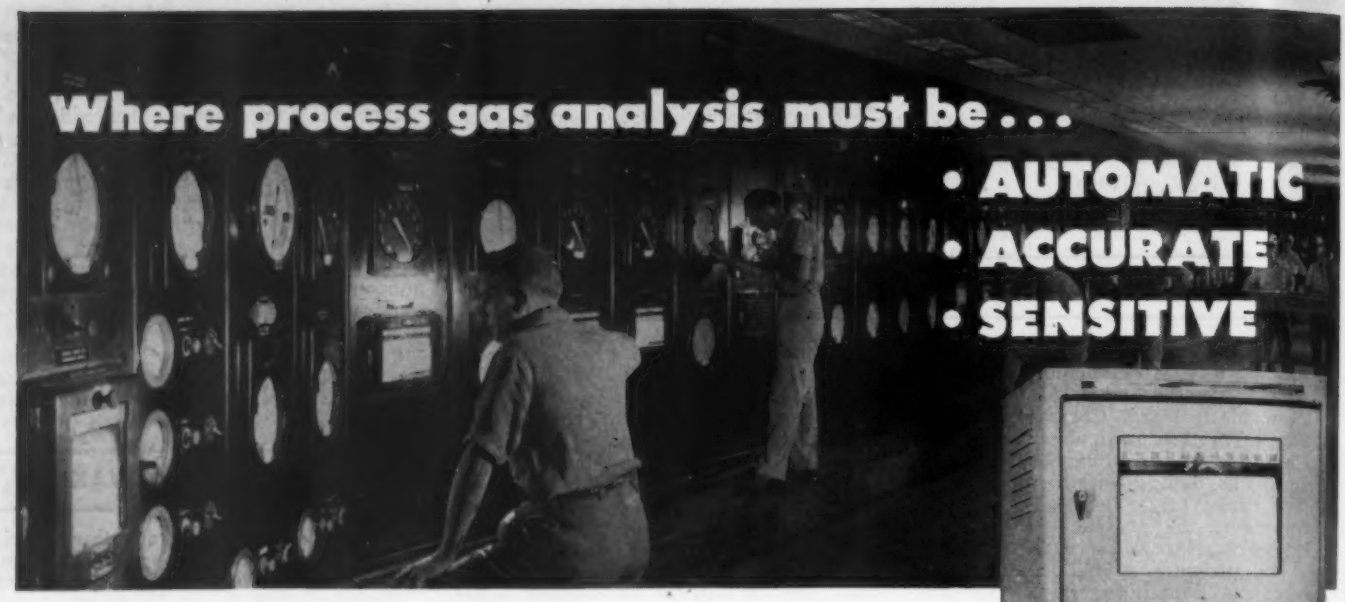


wherein each R is a saturated non-aromatic organic radical, and n and m are positive integers of at least 2 is dissolved in an aliphatic hydrocarbon with a Saybolt Universal viscosity of less than 50 seconds at 100° F. The resulting material is incorporated into a hydrocarbon lubricating oil, preferably by passing a gas through this oil for agitation. If the vehicle for the silicon compound is a relatively low boiling hydrocarbon, it can be removed after the incorporation of the silicon compound in the lubricant. The silicon compound, e.g. polymeric dimethyl silicone, can, for example, be used in an amount of 0.0001-0.5 percent by weight of the lubricating oil. Foaming of this oil is substantially reduced by the addition.

U.S.P. 2,435,206. Nonfoaming Compositions. G. H. B. Davis and J. C. Zimmer



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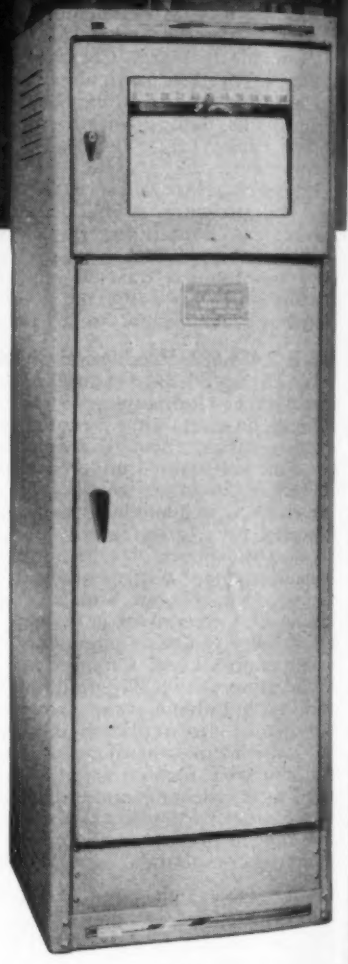

The Baird Associates infrared gas analyzer can be furnished in styles to suit various requirements, including remote indicating and recording, explosion-proof housings, and the cabinet style shown above.

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to Standard Oil Development Company.

The material according to this patent consists of a mineral lubricating oil, metallo-organic additives, such as barium tertiary octyl phenol sulfide, and 0.0001-10 percent by weight of highly fluorinated aliphatic hydrocarbons with 5-50 C atoms and composed of 60-80 percent F by weight, 20-37 percent C, and 0-3 percent H_2. These fluorinated hydrocarbons are oil-soluble and depress the excessive foaming properties otherwise caused in the oil by the metallo-organic additives.

U.S.P. 2,435,483. Preparation of Lubricating Oil. R. S. Vose to Sun Oil Company.

A hydrocarbon charge containing a substantial proportion of naphthenic constituents and less than about 10 percent aromatics, preferably a charge of this kind which boils above 400° F., is reacted with at least an equimolar quantity of an aliphatic alcohol with at least 8 C atoms per mol. under alkylating conditions and in the presence of a catalyst of the water-absorbing type, such as HF. Lubricating oils of high viscosity index are preferably obtained.

U.S.P. 2,435,542. Wax Acids Separation Process. F. G. Hess to Cities Service Oil Company.

Wax acids as obtained by the oxidation of hydrocarbon waxes, petrolatum, and the like are refined by treating a naphtha solution of the oxidation products with $ZnCl_2$. Naphtha soluble hydroxy acids are precipitated thereby and separated from the other oxidation products which can be further refined.

DESULFURIZATION

U.S.P. 2,434,868. Process for the Removal of Carbonyl Sulfide from a Hydrocarbon Fluid. G. E. Sample and W. B. Miller to Shell Development Company.

Carbonylsulfide (COS) is removed from a hydrocarbon fluid containing it by contacting this fluid with an aqueous alkaline solution in the presence of a water soluble aluminate, zincate, or chromite or mixtures thereof. Sodium aluminate is an example of such useful salts. The transformation of COS to CO_2 and H_2S by alkaline solutions is substantially accelerated by the presence of the soluble aluminate or the like.

PETROCHEMICALS

U.S.P. 2,434,313. Manufacture of Compounds Containing Carbonyl Groups. H. Dreyfus.

A mixture of one mol of a diolefin, such as 1:3-butadiene or 1:5-hexadiene, and at least 2 mols of formaldehyde is heated to a temperature of 50-80° C. The reaction mixture is cooled and the dialdehyde formed is separated therefrom. Adipic or suberic dialdehyde can, for example, be obtained.

U.S.P. 2,434,289. Manufacture of Hydrocarbons. L. Schmerling to Universal Oil Products Company.

An aliphatic mono-olefin is reacted with a monohaloalkane in the presence of a metal halide catalyst of the Friedel-Crafts type at a pressure sufficient to keep a substantial proportion of the re-

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actants in liquid state and at a temperature of —30° to 125° C. The resulting higher boiling monohalogenated saturated hydrocarbons are dehalogenated, for example by zinc and propanol to form the corresponding saturated paraffin or naphthenic hydrocarbons. Thus, 3-methyl-3-ethyl pentane can be produced from ethylene and sec.-butyl bromide, 3,3-dimethylpentane from ethylene and tert.-amyl bromide, and 1,1-diethylcyclohexane from ethylene and bromocyclohexane.

U.S.P. 2,434,403. Process for Producing Cyclopentadiene. F. E. Frey to Phillips Petroleum Company.

Propylene is continuously introduced into one end of a cylindrical turbulence chamber of relatively large cross-section at a high linear velocity while streams of acetylene are injected tangentially at a plurality of longitudinally distant points of this chamber at high velocity, the molar ratio of propylene to acetylene introduced being at least 9:1. Vigorous turbulent circulation of the reaction mixture is maintained in the chamber which is heated to 1200-1500° F. The treated mixture is continuously withdrawn from the opposite end of the chamber after an average residence time of 0.3-0.75 seconds. Cyclopentadiene is recovered from the cooled effluent of the process.

U.S.P. 2,434,850. Process for Production of Formaldehyde. C. A. Hochwalt, C. J. Stehman and R. W. Sudhoff to Monsanto Chemical Company.

A preheated gas mixture containing normally gaseous aliphatic hydrocarbons is oxidized in contact with surfaces composed of the heat reaction product of phosphoric acid on iron. The oxidizing gases can contain nitric oxide in catalytic amount. The contact surfaces can consist of phosphate, pyrophosphate, or metaphosphate of iron. Formaldephye is formed. The contact surfaces inhibit oxidation of the hydrocarbons or their first oxidation products to CO_2 and H_2O.

U.S.P. 2,434,888. Hydrogen Bromide-Catalyzed Oxidation. F. F. Rust and J. H. Raley to Shell Development Company.

A vaporous mixture comprising oxygen and a saturated aliphatic hydrocarbon containing a tertiary C atom, such as isobutane, is reacted in the presence of HBr at an elevated temperature below that at which spontaneous combustion of the mixture occurs, e.g. at 100-200° C., in a reaction vessel the inner glass surfaces of which are coated with an oxide of boron. Peroxidic compounds, such as tert. butyl hydroperoxide, di(tert. butyl-)peroxide and mixtures thereof are obtained. The use of a reaction vessel with inner glass surfaces treated with an oxide of boron has an activating effect on the catalytic reaction and high yields of the desired products are obtained. The boron oxide coating of the vessel can be reactivated by contacting its surface with an aqueous boric acid solution and subjecting the thus contacted surface to an elevated temperature.

U.S.P. 2,435,245. Stabilized Polymers of Ethylene. D. E. Strain to E. I. du Pont de Nemours & Company.

The material according to this patent consists of a normally solid homopolymer of ethylene and 0.001-5.0 percent by weight of diphenylamine-acetone con-

One of a series of messages to help you increase your understanding of business paper advertising, and its effect on your business.

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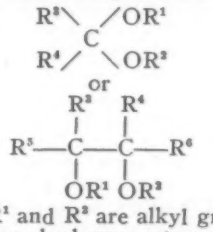
densation product. The ethylene polymer is stabilized against degradation by the additive.

U.S.P. 2,435,403. Cyclic Alcohols and their Preparation. R. C. Morris and T. W. Evans to Shell Development Company.

Butadiene or another conjugated diolefin is reacted with methyl vinyl ketone in the vapor phase to form a cyclohexenyl methyl ketone. This product is hydrogenated to a cyclohexyl methyl carbinol.

U.S.P. 2,435,524. Making Acetylenic Alcohols and Intermediate Formed therein. C. Weizmann to Polymerizable Products, Ltd.

Acetylenic hydrocarbons are interacted with aldehydes or ketones at temperatures not substantially above 0° C. in the presence of a suspension of a fluent reaction vehicle comprising a finely divided alkali metal hydroxide with a substantially larger amount of a water insoluble organic liquid containing in its structure two ether-forming oxygen atoms and having the formula



or

in which R^1 and R^2 are alkyl groups and R^3 to R^6 are hydrogen atoms or alkyl groups. This organic liquid has a b.p. substantially above 100° C. A chemically saturated acetal, a ketal, or a dialkyl ether of an ethylene glycol can, for example, be employed as such liquid. The activity of KOH, or the like, in promoting the reaction of acetylenes with aldehydes or ketones is markedly increased by the use of such organic solvents. Acetylenic alcohols are obtained.

U.S.P. 2,435,544. Method of Nitrating Organic Compounds. V. R. Kokatnur to Autoxygen, Inc.

Hydrocarbons such as benzene, toluene, or naphthalene, are mixed with a nitrating agent, e.g. HNO_3, and a water immiscible inert diluent, e.g. a saturated petroleum hbydrocarbon with a b.p. of 100-120° C., capable of forming an azeotrope with water. The amount of the diluent shall be sufficient to remove any water present in the mixture and also any water evolved by the reaction, and the azeotrope shall be vaporizable at the temperature and the pressure conditions of the reaction. m-Dinitrobenzene, trinitrotoluene, and nitronaphthalene can, for example, be produced by this process.

U.S.P. 2,435,545. Process for the Production of Mercaptans. J. P. Lyon to Phillips Petroleum Company.

A mixture of olefins and H_2S is passed over a silica gel catalyst promoted by an oxide of a metal of groups IIIB and IVA of the periodic system under mercaptan-forming conditions. Constant conditions of pressure, flow rate and concentration of reactants are maintained and the temperature is gradually increased during the reaction period to maintain a constant mercaptan-sulfur content in the effluent. Optimum selectivity of the catalyst toward the formation of mercaptans corresponding to the olefins charged is thereby obtained while the formation of lower-boiling mercaptans is excluded.

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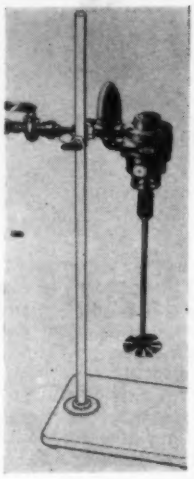
1—Air Motor Stirrer



The Emil Greiner Company, 161 Sixth Avenue, New York City, has developed a revolutionary new type of air motor stirrer, the "Mixaire," that achieves variable speed from 0 to 6000 rpm, and can replace electric motors in 0 to 1/6 hp range. This new device cannot explode under any conditions, does not generate sparks, and more than meets all safety requirements. Recommended for pressures below 100 pounds, the Mixaire's variable speed can be instantly increased or decreased by adjusting the valve on the air supply. Also, with four vanes acting in each revolution, the Mixaire will start from any position, and run at moderate speeds as low as 100 rpm without stalling. Among the Mixaire's other advantages are compact size (only 2½ inches in diameter) and weight of only 3 pounds. The Mixaire is supplied complete with pulley and chuck for shafts up to ¼-inch diameter, nickel-plated stirrer, 9-inch shaft, and 2-inch diameter propeller.

2—"Artistic" Sprinkler

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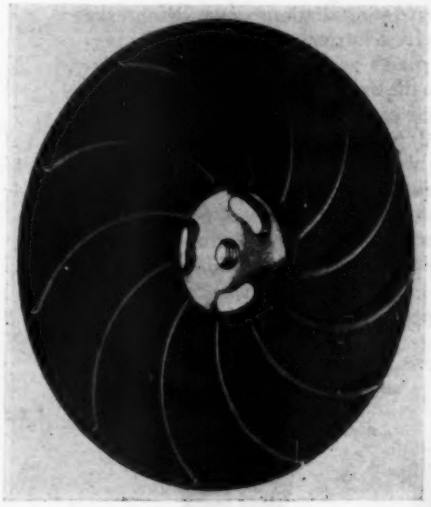
Quartzoid Ceiling Sprinkler

3—Cooling Tower

To meet the demand for a "small natural draft cooling tower that embodies the engineering design, balanced spray system and predetermined performance only previously available in large towers," The Marley Company, Inc., Kansas City 15, Kan., has announced the new Series 100 tower. Features are lumber fabricated from heavy California heart redwood, all hardware galvanized, machined louver posts without troublesome nails, bolts, or backup strips, slip-fit louvers that are readily removable and frame and girts bolted two ways to give maximum strength. Marley patented bronze spray nozzles are used to give "maximum efficiency at the specified gallonage." Ask for Bulletin 100-48.

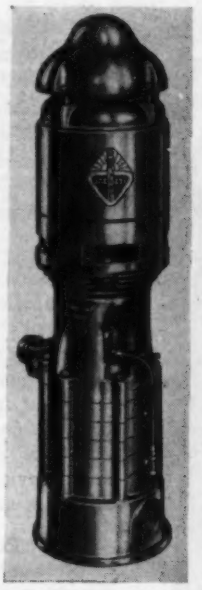


Sanding Disc Backing Pad

4—Sanding Disc Backing Pad

A new Spiralcool backing pad for disc sanders, produced by The Pratt Manufacturing Company, Inc., 614 East Madison, Fairmount, Ind., is made of layers of rubber reinforced with fabric and securely bonded to a slotted iron hub. The slots in the hub forces air through to the back of the sanding disc, keeping it cool. The forced draft action through the spirals on the pad tends to keep the abrasives clean. The pads, being made of rubber, allow greater flexibility in working various shapes and contours without the danger of gouging the work, and provides greater contact area of the disc.

5—Boiler Feed Pumps

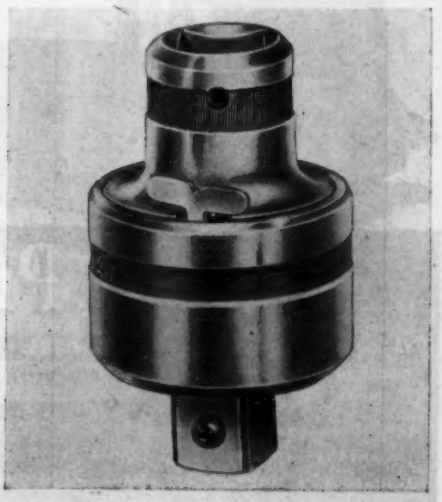


Jacuzzi Bros., Inc., Richmond, Calif., have added to their line of pumps a series of boiler feed pumps especially designed for pumping hot water and for other installations requiring high water pressure. These multi-staged centrifugal pumps are said to be the first boiler feed pumps to be produced in a vertical design. This cuts to a minimum floor space requirements and does away with the need for special foundations or flexible couplings. The new pumps have only one moving part and no close clearances. Standard units vary in size from ½ to 20 hp. and in addition the firm will make custom-built pumps for specific requirements.

6—Socket Wrench Attachment

—Socket Wrench Attachment .. .. ..

Any ½-inch drive socket wrench handle or attaehment can be converted into a ratcheting device with a new tool, known as the "Ratchetor," announced by Plomb Tool Company, Los Angeles 54. The tool has a ½-inch square plug with ball-check for holding a socket wrench, a ½-inch square opening for

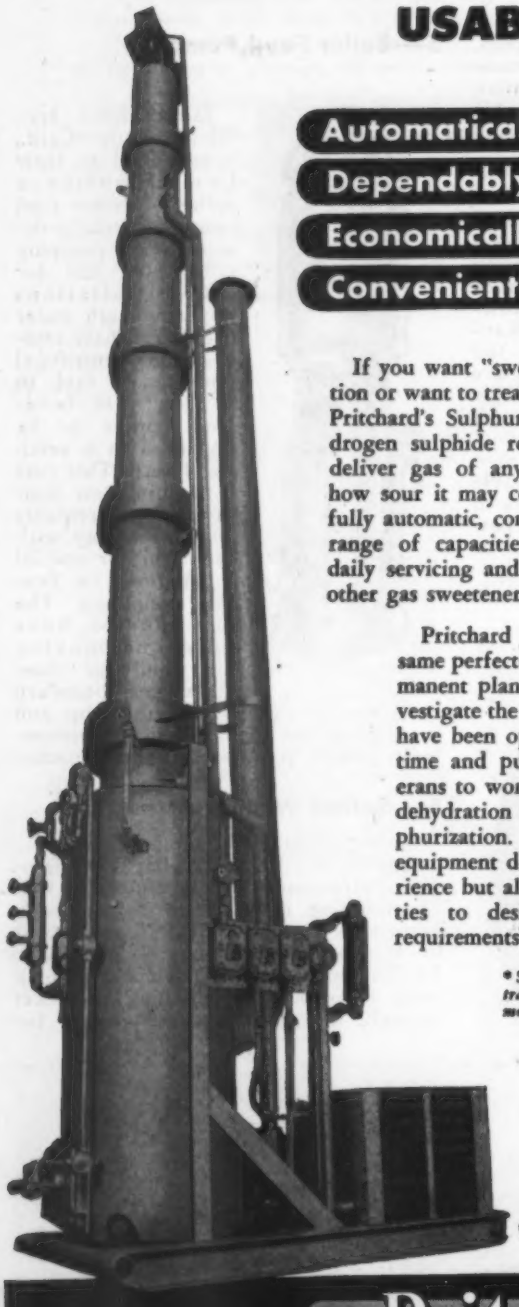


Socket Wrench Attachment

PRITCHARD'S New Gas Treater

the SULPHURSORBER*

Converts SOUR GAS into USABLE FUEL



Automatically
Dependably
Economically
Conveniently
} in any quantities, anywhere

If you want "sweet" gas at an isolated field location or want to treat large volumes at a central plant, Pritchard's Sulphursorber is the answer. These hydrogen sulphide removal units of the amine type deliver gas of any specified purity, regardless of how sour it may come from the well. Available in fully automatic, compact, *portable* design in a wide range of capacities, Sulphursorbers eliminate the daily servicing and hauled-in materials required by other gas sweeteners.

Pritchard installations employing the same perfected principles in large-scale permanent plants are equally economical. Investigate the many Pritchard projects which have been operating over long periods of time and put this qualified team of veterans to work on your own gas problems, dehydration and cleaning as well as desulphurization. Pritchard offers not only the equipment developed from thorough experience but also the organization and facilities to design and construct for any requirements.

* SULPHURSORBER *is a registered trade name for gas purification equipment produced by J. F. Pritchard & Co.*

J.F. P Co.

EQUIPMENT DIVISION
908 Grand Ave. • Kansas City 6, Mo.

See Sweets Files, Chemical Engineering Catalog, Refinery Catalog, etc., for additional information on Pritchard engineering and construction services and equipment.

J.F. Pritchard & Co.
ENGINEERS • CONSTRUCTORS • MANUFACTURERS
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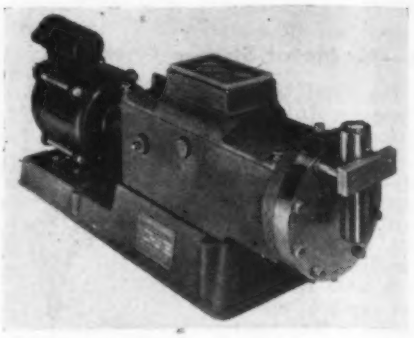
LITTLE ROCK • RICHMOND • OMAHA • MEMPHIS • BALTIMORE • CEDAR RAPIDS • AMARILLO
PHILADELPHIA • NEW ORLEANS • OKLAHOMA CITY

insertion of a handle or attachment, and a reversible ratcheting mechanism. It is built for high torque-strength—yet compact for ample clearance. A ball-type thrust bearing and light spring tension is said to result in very little back drag.

7—Proportioning Pumps

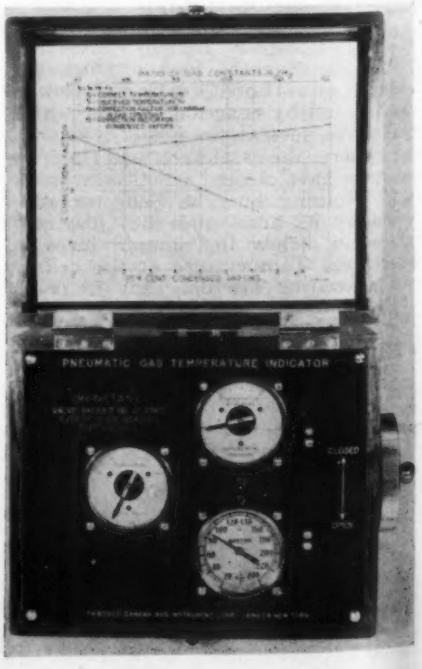
Lapp Insulator Co., Inc., 730 Maple Street, Le Roy, N. Y., has announced a new line of Pulsafeeder proportioning pumps. Designated as CP models, the new series is said to provide greater capacities, higher pressure operation and



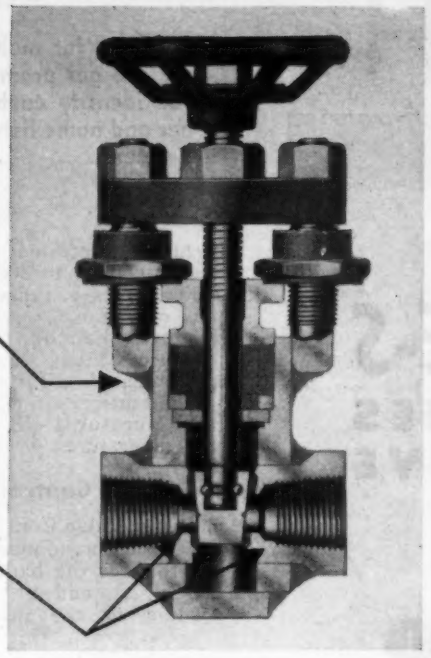
rugged assembly for accurate feeding. The pump, a combination piston-diaphragm unit, provides the accuracy of positive piston displacement and measured feed, while at the same time incorporating the diaphragm pump advantages of elimination of stuffing box, and isolation of chemical from pump parts. They are micro-adjustable over the entire capacity range while the pump is in operation. Four sizes are available ranging up to 330 gph. Ask for Bulletin 262.

8—Gas Temperature Indicator

Fairchild Camera & Instrument Corporation, 88-06 Van Wyck Blvd., Jamaica 1, N. Y., has announced Unit 263 laboratory gas temperature indicator for measuring gas temperatures to 5000° F. Insertion of a probe into the gas stream and operation of a single control valve on the instrument case is all that is required to obtain gas temperature meas-



Gas Temperature Indicator



Three joints less!

EACH of these three joints in the conventional gate valve design is a potential source of leakage. *There are no joints in the* Hancock WELDVALVE!

There is no bonnet joint in the Hancock WELDVALVE, because it has no bonnet.

There are no seat insert joints, because the super-hard cobalt-chromium faced seats are integral with the valve body and never need to be replaced. Excess handwheel pressure to align distorted seating surfaces is not necessary for WELDVALVE tightness, because the alignment of the angle of the seats with the "500 Brinell" stainless steel wedge is finished to *exact* dimensions before assembly.

Designed for 800# at 750° F or 2000# at 100° F. Sizes ¼" to 2" inclusive.

Stocked and sold by leading Distributors everywhere. Write to them or to us for full information.

1877 HANCOCK U.S.A.

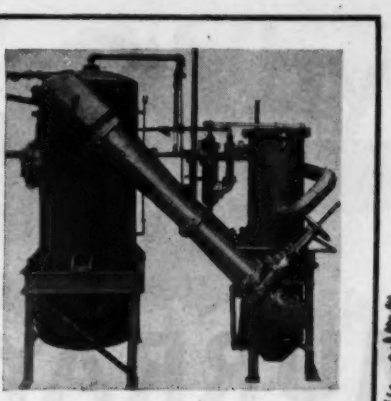
MANNING MAXWELL & MOORE INC. M&M TRADE MARK

HANCOCK *Valves*

A Product of

MANNING, MAXWELL & MOORE, INC.
WATERTOWN 72, MASSACHUSETTS

Makers of Hancock Valves, Ashcroft Gauges, Consolidated Safety and Relief Valves and 'American' Industrial Instruments. Builders of 'Shaw-Box' Cranes, 'Budgit' and 'Load Lifter' Hoists and other lifting specialties.



ORDER YOUR CHILLVACTOR NOW

For Chilled Water Next Summer

Every year when the warmest weather comes in July and August, we have several frantic appeals from production men in petroleum and process industries for CHILL-VACTOR equipment to maintain proper temperatures in certain solutions, in chilled water for air conditioning, for chilling various reaction vessels, and for special cooling requirements.

The Croll-Reynolds CHILL-VACTOR is one of the most dependable and efficient units available for cooling requirements in the range above 32° F. While there are a great many outstanding advantages, there are a few limitations. In recent years the most serious one seems to be the fact that these units are not made in quantity and available for immediate shipment. While they are mostly made from standard parts, each unit is custom built to a certain extent in order to provide maximum dependability and efficiency under individual operating conditions.

Under present manufacturing schedules, it requires approximately three months to design, build and ship a complete CHILL-VACTOR unit. This means that **immediate** action is advisable for those who want to be ready with new cooling capacity for the next warm weather season.

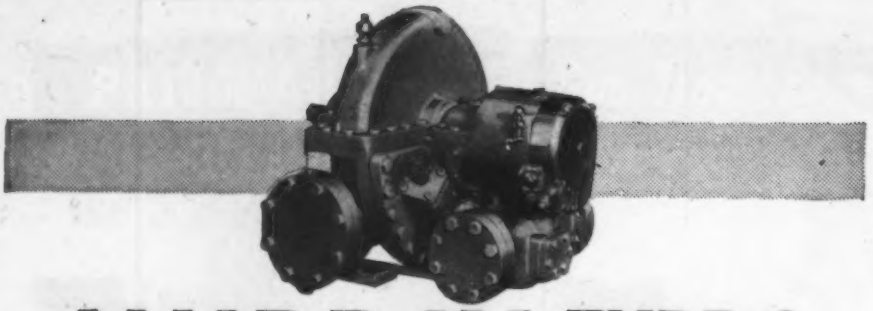
The Croll-Reynolds CHILL-VACTOR has no moving parts other than a standard chilled water circulating pump. Water is the only refrigerant. Maintenance and repair costs are therefore practically nil. Operating costs are usually less than for other types of refrigerating equipment where ample condenser water is available. Inquiries on this type of equipment will receive the careful attention of engineers with many years specialized experience in this field.

CROLL C R REYNOLDS

CROLL-REYNOLDS CO.

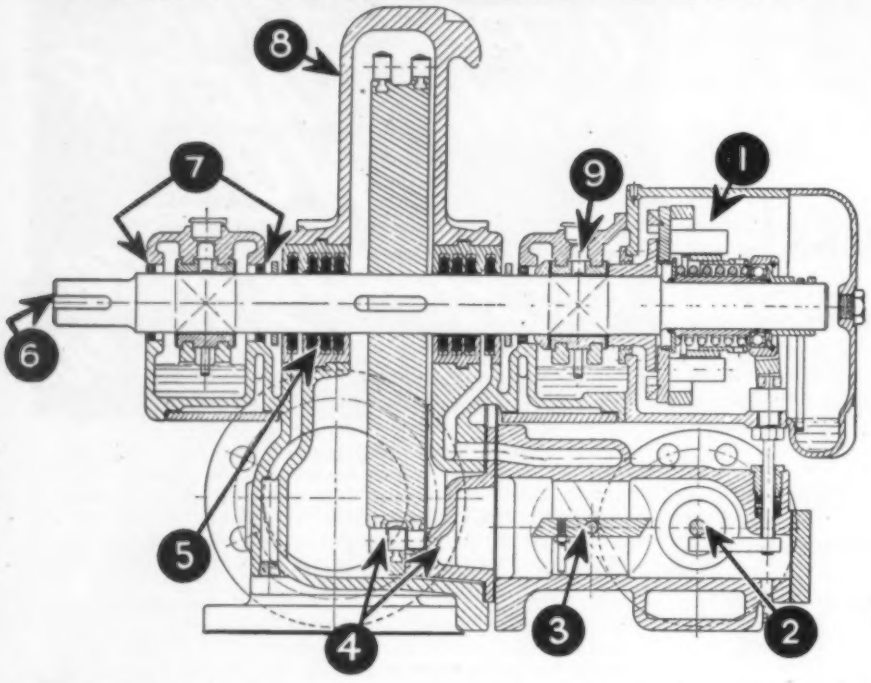
17 JOHN STREET, NEW YORK 7, N. Y.

CHILL VACTORS STEAM JET EVACTORS CONDENSING EQUIPMENT



MURRAY *TYPE-S*

SINGLE STAGE TURBINES FOR MECHANICAL DRIVE



1. Mechanical constant speed governor completely enclosed and fully lubricated. Heavy cast housing, removable for inspection. Provision for tachometer reading, by removing end plug.
2. Double seated balanced stainless steel governor valve, with stainless steel renewable seats. Corrosion resisting steam strainer ahead of control valves.
3. Separate emergency overspeed butterfly trip valve actuated by unbalanced weight type emergency overspeed governor.
4. Stainless steel blading and shroud band. Nozzles drilled and reamed in replaceable nozzle ring.
5. Carbon ring glands, stainless steel springs, corrosion resisting gland cases. Cast in leakoff reduces pipe fittings necessary.
6. Tapped hole in shaft end facilitates drawing on coupling.
7. Bearing seals effectively retain oil and exclude dirt and moisture.
8. Casing split at centerline allowing inspection without disturbing alignment or steam connections.
9. Sleeve type babbitt lined bearings with split oil rings, easily removable. Babbitt thrust faces against collars. Inspection covers and flush level glasses included. Water cooling always furnished.

FOR DEPENDABILITY, SPECIFY A MURRAY TURBINE

Bulletin T-121 gives complete details. Write for your copy today.

MURRAY IRON WORKS COMPANY

BURLINGTON, IOWA

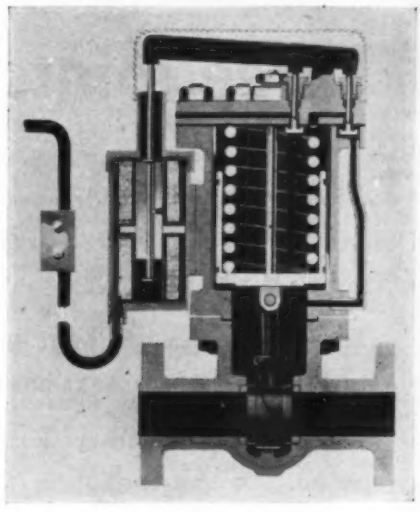
Builders of Steam Power Equipment for Three Quarters of a Century

Requests for manufacturers' literature get prompt replies when you identify each item by number and name listed in Petroleum Refiner.

urements. Accuracies of 1 percent are guaranteed up to 2000° F. In the 2000-5000 degrees F. range, reproducibility of plus or minus 2 percent is guaranteed. The basic Fairchild pneumatic gas temperature system, the heart of this new instrument, is a form of gas thermometer and utilizes the density of the gas being measured as a direct indication of temperature.

9—Valve Controls

Shafer Valve Company, Inc., of Mansfield, Ohio, announce a full line of valve controls for the handling of steam, water, oil, gas and air. These controls, according to the manufacturer, extend



valve life, provide efficient control for high or low pressures and temperatures and may be set for fast or slow operation. Both electrically and manually operated controls are available in a full line of sizes.

10—Lightweight Torch

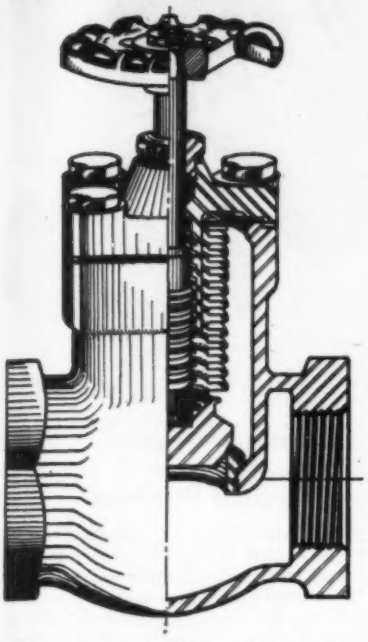
Victor Equipment Company, 844 Folsom Street, San Francisco, has announced its model 3401 "four-in-one" torch for light brazing or soldering operations. It can be used with any practical fuel gas, in combination with compressed air or oxygen. Needle valves are conveniently located for finger tip control and the parallel grip may be easily clamped in a small vise or special fixtures. It is furnished with three tips and a 4-way wrench.

11—Welding Gloves

Air Reduction Sales Company, 60 East 42nd Street, New York 17, has added cowhide welding gloves to its line of welding supplies and equipment. The gloves are made of medium weight, chrome tanned cowhide with a one piece leather back employing the wing-type thumb construction and fabric-lined cuff. Exposed seams across the back of the glove have been eliminated thereby minimizing the possibility of cuff separation.

12—Packless Valve

A new type packless valve announced by Fulton Sylphon Division of Robertshaw-Fulton Controls Company, Knoxville, Tenn., is termed the "No. 3000 series," and incorporates a seamless



metal Sylphon bellows which forms a leakless connection between the valve disc and the valve body. Thus no maintenance is required to keep the stem seal tight against leakage. The bodies come in bronze, cast iron, cast steel, Monel metal, or stainless steel. The Sylphon bellows is available in either stainless steel or Monel, and can also be supplied in a variety of other metals including phosphor bronze and brass. The valves are made in sizes ¼-inch up to three inches with screwed, welded or flanged ends.

13—Fire Protection

American-LaFrance-Foamite Corporation, Elmira, N. Y., has developed a completely new carbon dioxide fire extinguishing system for Class B and C risks requiring localized or total flooding and for rotating units requiring a sustained discharge of CO_2. Automatic sounding of alarm and then automatic cutoff of power units, window and door closings precede discharge of the chemical. The unit is flexible, it being possible to add cylinders should a risk be increased.

14—Face Shield

American Optical Company, Southbridge, Mass., has announced a new plastic face shield specially designed to be worn in combination with the company's R-1000 and R-2000 respirators. Light in weight, durable and easy to attach to the respirators, this shield pro-

For free copies of manufacturers' literature or more information about new equipment featured in Petroleum Refiner, simply identify the product by name and number.

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Equipment . . . . . or . . . . . Other Classification—**please identify**

☐ Manufacturer
☐ Supply Company
☐ _____

Return to P. O. Box 2608 Houston 1, Texas

Face Shield for Respirator

tects the eyes and major part of the face against the impact of foreign particles striking from the front. Outer edges of the shield are protected by a vinyl plastic binding. The shield may be worn over regular prescription glasses.

15—Roof Fastener

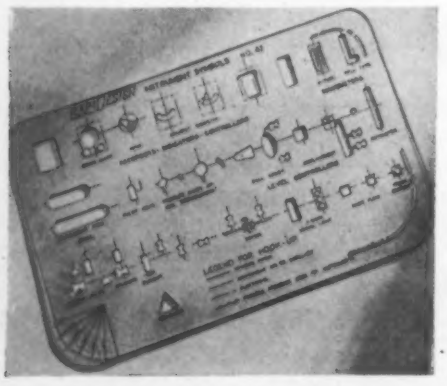
New type roofing fasteners have been placed on the market by Keasbey & Mattison Company, Ambler, Pa. The customary expense of scaffolding and workmen beneath the roof are entirely eliminated because the whole installation is completed from the roof surface. In addition, cementing bolt heads is not necessary because of lead covering tabs which permanently weatherproof each "Top-Side" fastener, the manufacturer points out. It also provides for movement between steel frame and the asbestos corrugated roofing.

16—Electrode

A new Airco No. 387 all-position arc welding electrode for welding of mild steel has been developed by Air Reduction Sales Company, Department 1815P, 60 East 42nd Street, New York 17. Of the E 6012 class, the bead is smooth and of proper contour, "excellent operating characteristics" are obtained on ac or dc and it operates most satisfactorily in all positions, especially when welding vertically down. This electrode is available in ⅛, 5/32, 3/16 and ¼-inch diameters.

17—Template Guide

Rapidesign, Inc., Box 592, Glendale, Calif., announces the new No. 47 instrument symbols template, particularly adapted to layouts where liquid level, differential pressures and flow instruments are used. Schematic diagrams of recorders, indicators, controllers, valves manometers and liquid level controllers are included. The template, 5 x 8 inches, is made of clear, .040 mathematical-quality plastic with printing on reverse side. Cut outs are beveled and the template has a 90 degree protractor and a dotted line radius guide.



Instrument Symbols Template

18—Cryostat

Experiments at —456° F. are possible by means of the Collins helium cryostat. After Dr. S. C. Collins' (Massachusetts Institute of Technology) first Cryostat was completed in 1946, Arthur D. Little, Inc., industrial research organization, Cambridge 42, Mass., developed a design which could be operated by laboratory technicians. The Cryostat liquefies helium gas, just as the household refrigerator liquefies Freon, sulfur dioxide or other refrigerant. The cold space of the Collins Helium Cryostat and part of the refrigerating equipment is contained within a steel vacuum jacket 12 inches in diameter and about five feet tall, which is housed within a cabinet measuring about three feet by two. Other equipment includes a ten-horsepower compressor and a low-pressure gas holder. The complete unit weighs about 2800 pounds and occupies less than 50 square feet of floor space.

19—Tally Board

An electrically illuminated tally board called the "Tally-Lite," which throws a clear light on tally sheets for speeding checking and entering where there is little or no light, has been announced

by Clyne Manufacturing Company, 2619 Collerain Avenue, Cincinnati 14, Ohio. The unit is compact, sturdy and fits comfortably in one hand, leaving the other hand free for entries. The entire unit with three standard flashlight batteries weighs only 24 ounces.

20—Laboratory Kit

Fischer & Porter, Hatboro, Pa., have announced a new laboratory kit (rotameter type), "a versatile and inexpensive flow measurement set to meet the varied flow measurement requirement of the research lab." It consists of a unique metering tube holder with base plate and hose connectors, four readily interchangeable tubes and six metering floats complete with calibration data. Capacities from 0.065 to 2200 cc/mm of water and 5.0 to 36,800 cc/min of air are covered.

Laboratory Kit (Rotameter Type)

21—Recording Wattmeter

A new photoelectric recording wattmeter designed for such applications as analyzing welding circuits and measuring industrial heating load, motor-starting load, and power surge has been announced by the meter and instrument divisions of General Electric Company, Schenectady 5, N. Y. Consisting of a high-speed photoelectric recorder and a newly designed wattmeter measuring unit, the new instrument can be used to measure dc power or ac power up to 10,000 cycles per second. The wattmeter, which has a linear scale, is available for portable use or for semiflush mounting. Chart speeds of a half inch per hour up to 72 inches per minute can be obtained by using different combinations of chart carriages and rate gears.

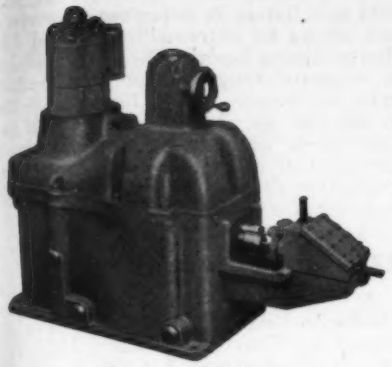
22—Fluid Catalyst Filter

Micro Metallic Corporation, 193 Bradford Street, Brooklyn 7, announces availability of filters designed for use in fluid catalyst pilot plant installations. The filters are fabricated from "Micro Metallic" porous stainless steel sheet ⅛-inch thick. The pressure drop through the filter elements is very low and the filter elements appear not to plug up with time even when used with the finest catalyst, the manufacturer reports. The filter cake is discharged instantly from the whole surface upon application of the blowback pressure.

23—Chemical Pump

The "constametric," an original type of chemical pump, has been announced by Milton Roy Company, 1315 East Mermaid Avenue, Philadelphia 18. It is a piston type pump developed especially to handle liquids in precisely controlled volume, to deliver at a constant rate of flow (without pulsation), and to pump in

When writing the manufacturer of new equipment for literature or more information identify his product by name and number listed in Petroleum Refiner.



"Constametric" Pump

minute quantities, down to as little as 75 ml. per hour as a maximum for an individual pump.

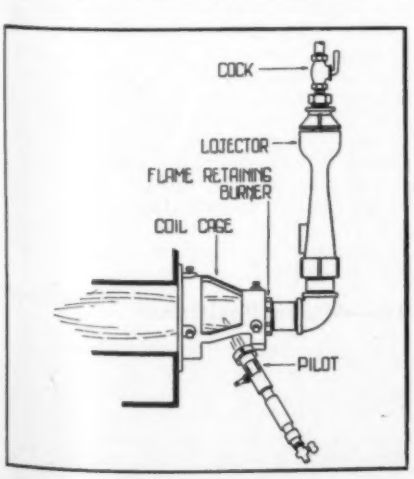
The company calls attention to the flexibility of the unit, the use of interchangeable gears to vary capacity, and the availability of various corrosion resistant materials for liquid ends. The pump will be in production in six to eight months.

24—Torque Control

Eclipse-Pioneer Division, Bendix Aviation Corporation, Teterboro, N. J., has announced development of a remote torque control system, that provides an electrical means for either manually or automatically transforming a very weak motivating force into control of a predetermined operation at a remote location with a high degree of accuracy. The torque system comprises three units: a transmitter Autosyn (synchro), a torque unit and an amplifier. The position of the Autosyn rotor shaft determines the position of the torque unit output shaft. The Autosyn shaft can be set manually or by elements sensitive to pressure, temperature, humidity, volume, etc. The amplifier supplies the controlled power required to operate the torque unit. Ask for Bulletin 711-21.

25—Gas Burner Units

Newly-engineered immersion heating burner assemblies are announced by the industrial division of Bryant Heater Company, 1020 London Road, Cleveland 10, Ohio. The assemblies involve scientifically-designed spacer cages which "insure the entrainment of the required amounts of secondary combustion air in gas-firing immersed tubes for heating water, oils and other liquid solutions." The "coil cages" are furnished separately, or in ready-for-installation assemblies, complete with mixers and associated equipment. Ask for Bulletin 1C-1.



Typical Atmospheric Burner Immersion Coil Combination

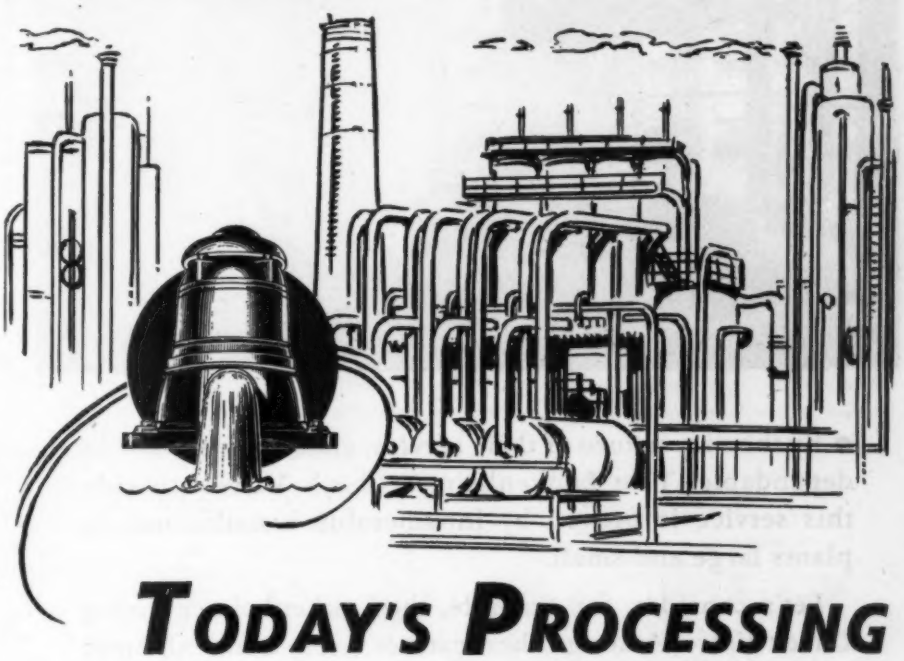
When writing the manufacturer of new equipment for literature or more information identify his product by name and number listed in Petroleum Refiner.

26—Hand Power Cutters

H. K. Porter, Inc., Somerville, Mass., has redesigned its line of hand power cutters. Improvements include curved toe-in handles which decrease armspread and wrist bend making an easier and stronger pull, broad rounded palm-fit grip which reduces fatigue, no protruding screws or bolts to catch, metal safety handle stop, 30 percent greater tool strength in sections and handles, jaws plainly marked showing capacity and type of metal safety cut, and red color finish.

27—Vertical Turbine Fire Pump

Vertical turbine fire pumps, manufactured by Peerless Pump Division of the Food Machinery Corporation, 301 West Avenue 26, Los Angeles 31, has received full approval by Underwriters' Laboratories and other agencies. This type of vertical turbine fire pump is said to be



TODAY'S PROCESSING

DEMANDS LOW COST WATER SUPPLY

Refiners faced with many added processing operations are in need of more—and lower cost water. Old, low efficiency systems are eating heavily into profits and production is often being retarded. The way to change the situation is to install a Layne well water system and thus obtain larger quantities of water at lower cost. Layne well water systems are world famous for high efficiency, low operation cost, dependable service and long life. Once installed according to Layne standards, there is little or no upkeep cost. For literature, address Layne & Bowler, Inc., General Offices Memphis 8, Tenn.

√

Layne well water systems are designed and built in capacities from several thousand up to many million gallons daily. Installation may be of the all weather type requiring no housing. Controls are completely automatic. Flow may be direct into the mains or to elevated tanks.

LAYNE WELL WATER SYSTEMS

● By the very nature of their service, check valves *must be* dependable. That Lunkenheimer Check Valves provide this service is proved by innumerable installations, in plants large and small.

Let's consider, for example, the Lunkenheimer Swing Check Valve. Among other features, every Lunkenheimer Swing Check Valve is equipped with *two* renewable side plugs. Normal carrier pin wear can be compensated for easily. Recognized by leading valve users as remarkably trouble-free, Lunkenheimer Check Valves provide positive reverse-flow protection over a much longer, lower-cost service life.

Your Lunkenheimer Distributor has a stock of Check Valves for your needs. Call on him for your original equipment, replacement, and repair valve requirements.

. . . More than 96 types of LUNKENHEIMER *Check Valves*

Available in sizes and materials to meet every service requirement.

From small Bronze 125 lb. S.P. to massive steel patterns for pressures up to 2500 lb. S.P.

Lunkenheimer Check Valves are described and illustrated in Circular Number 588. Ask your Distributor for a copy or write direct.

ESTABLISHED 1862
THE LUNKENHEIMER CO.
"QUALITY"
CINCINNATI 14, OHIO. U. S. A.
NEW YORK 13 CHICAGO 6
BOSTON 10 PHILADELPHIA 34
EXPORT DEPT. 318-322 HUDSON ST., NEW YORK 13, N. Y.

virtually impervious to deterioration, permits installation in minimum floor space and allows an extremely simplified discharge piping layout. They are available in capacity ranges from 500 to 2000 gpm. to operate against heads from 240 to 285 feet. Buyers have a choice of not only oil or water lubrication but also a choice of drives, either electric or right-angle gear drive, or steam turbine drive. Ask for Bulletin B-152.

28—Fork Lift Accessory

A new fork lift truck accessory, the "Multiple Drum Carrier," has been designed by Trayner-Reinhart, 1627 Willow Street, Oakland 7, Calif. Towmotor Corporation, 1226 East 152nd Street, Cleveland 10, Ohio, has been licensed to manufacture the carrier under the trade name "Towmotor Multiple Drum Carrier," whereas Trayner-Reinhart will market it under the trade name of "Tray-Hart Multiple Drum Carrier." The accessory is said to provide safe, easy and fast handling of 1-2-3 or 4 drums, by means of specially-designed, hydraulically-operated gripping shoes, mechanical and gravity action providing a strong, positive grip. Combined with Towmotor's high-stacking ability, the multiple drum carirer also facilitates stacking of drums as well as loading operations.

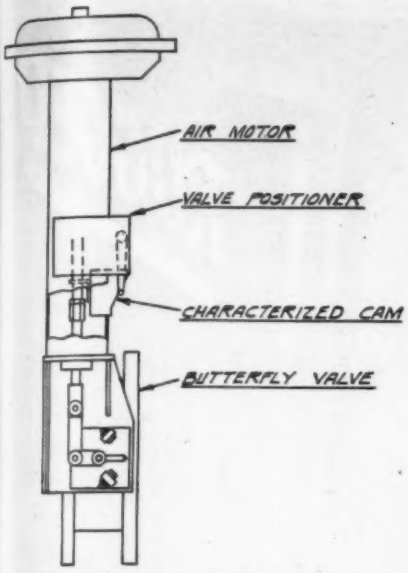
Multiple Drum Carrier

29—Lubricant and Compound

Masters anti-acid lubricant and Masters metallic compound have been announced by James K. Harbinson & Company, Inc., 70 West Chippewa Street, Buffalo 2, N. Y. The lubricant, when worked into packing fibers, is said to prolong the useful service life of the packing while the metallic compound is a thread dope which was "developed to protect threads in all types of service to prevent them from bonding or galling so that after extended service the nut may be removed without difficulty."

30—Valve Positioner

A cam-operated positioner is a key feature of a new diaphragm motor valve of the disc or butterfly type announced by R-S Products Corporation, Philadelphia 44. Excellent control of flow characteristics is said to be attainable by simply changing cams in the field on the cam-operated positioner, which, in effect, adds to the normal functions of the



Hook-Up of Cam-Operated Valve Positioner Installation.

disc type valve. The desirable characteristics of a straight line, semi-log curve (V-port), ratio plug valve, throttle plug valve, linear, or almost any other special characteristics to suit individual process requirements, are claimed within the range of the new positioner.

31—Programming Control

Fireye programming control type 24PJ8, manufactured and sold by Combustion Control Corp., 77 Broadway, Cambridge 42, Mass., is designed to provide automatic starting and programming control for commercial and industrial oil burning equipment. It is used

with photoelectric flame failure scanner type 45PH5 which protects the oil burning installation from the hazard of flame failure. The control automatically starts a burner in operation and programs a sequence of fuel valve opening, post-ignition time, priming, and scavenging. Bulletin CD4751 describes the equipment.

32—End Seamer

A new high-speed automatic end seamer for square and irregular-shaped cans has been announced by Lima-Hamilton Corporation, 60 East 42nd Street, New York 17.

Called the Hamilton-Kruse Model 150, this is a completely automatic four-head unit with four round rollers on each head. In operation, the cans remain stationary on the base plates during the seaming. It is equipped with automatic body and end feeds, a complete set of seaming parts, and its own motor and control. Its features also include low speed heads, exceptional ease of adjustment, and automatic lubrication.

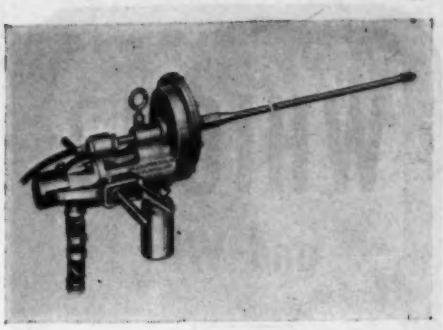
Requests for manufacturers' literature get prompt replies when you identify each item by number and name listed in Petroleum Refiner.

33—Electric Etcher

Anything made of steel, iron or their alloys, can be easily marked with a new electric etcher manufactured by Ideal Industries, Inc., Sycamore, Ill. Named the "Universal" model, the unit is inclosed in a streamlined, metal case for ready portability. When opened, the cover forms a work plate. The etching tool does not get hot because the point is so mounted in the handle that cooling air can circulate around it. Four etching heats are provided, 120, 240, 420 and 700 watts.

34—Tube Cleaner

A geared tube cleaner, (No. CC-325), has been developed by Airetool Manufacturing Company, 316 South Center Street, Springfield, Ohio, for removing extremely bad scale in heat exchanger and condenser tubes ⅜ to 2½ inches i.d.



New Geared Tube Cleaner

It is said to be particularly effective where conditions require low speed and extreme power. The motor operates up to a maximum of 800 rpm and will clean tubes even though completely closed with scale. Various types of drills are available for use with this cleaner. Ask for Bulletin 40.

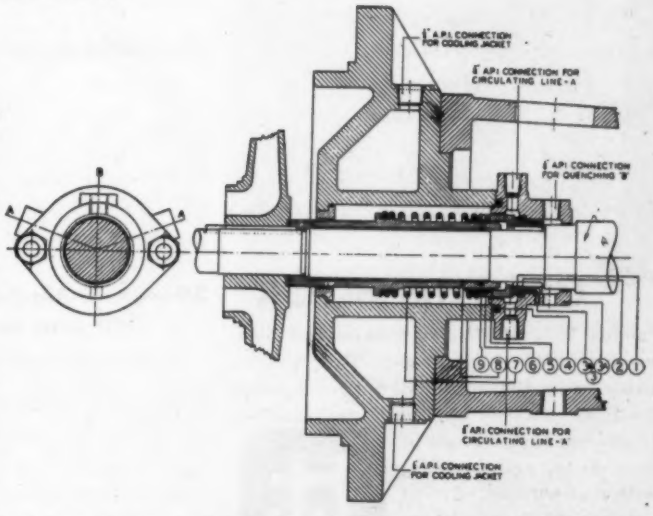
36—Recording Mass Spectrometer

A new recording mass spectrometer capable of measuring over the mass range from 1 to 350 and adjustable to any portion of that range, has been announced by General Electric Company, Schenectady, N. Y. This instrument incorporates the control and stability features which have made the mass spectrometer a practical method for the chemical analysis of gases and

35—Seal for Process Pumps

A new type of mechanical seal (Type RR) for process pumps has been announced by Flexibox Ltd., a member of the M. O. R. group of companies, 10 Dover Street, London, W. I., England.

The seal fits inside the conventional stuffing box and does not entail major alterations to the pump. The usual hard-faced shaft sleeve is replaced by a stainless steel sleeve (8) and a static seal (9) serves to prevent quenching water finding its way under the shaft sleeve (8).


¼" API CONNECTION FOR COOLING JACKET
¼" API CONNECTION FOR CIRCULATING LINE - A
¼" API CONNECTION FOR QUENCHING "B"
¼" API CONNECTION FOR CIRCULATING LINE - A
¼" API CONNECTION FOR COOLING JACKET

Seal Type RR for oil refining process pumps. 1–Stationary seal ring; 2–cover seal; 3–cover assembly; 3a–cover; 3b–pin; 4–joint ring; 5–rotary seal ring; 6–sleeve seal; 7–spring; 8–shaft sleeve; 9–shaft seal.

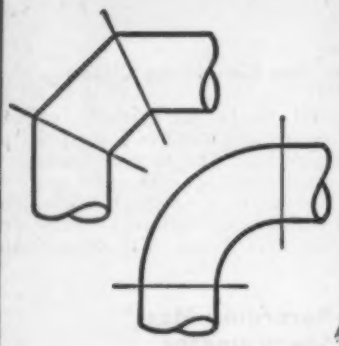
A unique feature of the RR seal is the patented spring drive. The stainless steel spring (7) is an interference fit on the thickened portion of the shaft sleeve (8) at one end and on the rotary seal ring (5) at the other end. In addition, the sense of winding of the coils of this spring is such that, in operation, it tends further to increase its grip. Consequently, the rotary seal ring (5) is driven positively yet resiliently and the static seal (6) is not subjected to frictional wear. It is claimed by the manufacturer that prolonged tests with various designs of seals have shown erratic behaviour to be due, in most cases, to side thrust transmitted to the rotary seal ring by driving pegs, screws, tongues, etc. The spring drive incorporated in the RR seal is said to be a perfect solution to this problem.

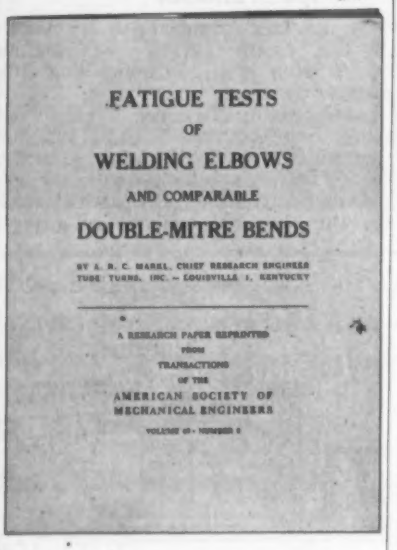
A stationary carbon seal ring (1), is arranged inside the stainless steel cover (3A). The static seal (2) is engaged by a peg (3B) which prevents its rotation.

Two tapped holes ("A") are connected to the discharge and suction sides of the pump respectively, thus providing effective circulation of the pumped liquor around the two seal rings to remove frictional heat and prevent the accumulation of sediments. A valve in one of these lines controls the amount of circulating liquor.

WHICH?

and Why





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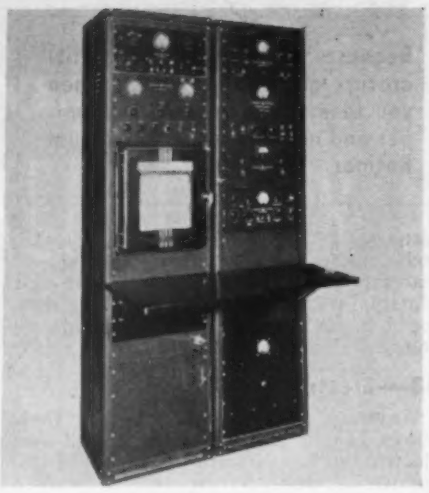

vapors. Applicable to general chemical analysis and isotope abundance measurements, the instrument is built in two main units. The control unit contains the electronic components and recorder, while the tube rack houses the spectrometer tube, vacuum pumps, gauges, sampling system, and additional components. Ask for GEA-4907.

37—Electronic Control

R-S Products Corporation, Wayne Junction, Philadelphia 44, has published a detailed catalog (No. 27) on its new Leveltronic relay, designed for basic applications in the control of liquid level, interface, pressure, and temperature.

38—pH Control

Milton Roy Company, 1315 East Mermaid Avenue, Chestnut Hill, Philadelphia 18, has reprinted "Automatic pH Control in Water and Industrial Waste Treatment" by Robert T. Sheen, the company's consulting engineer, and has announced that this Technical Paper 55 is available upon request.

39—Heat-Medium Salt and Boiler

Bethlehem Foundry & Machine Company, Bethlehem, Pa., has available a folder just published which contains technical details and performance data of the Beth-Tec unit, a specially designed boiler for utilization of a new heat-medium salt.

40—Blowers

The blower department of Elliott Company, Jeannette, Pa., has issued Bulletin P-7, on its Type P single-stage centrifugal blowers using open type radial bladed impellers. Shown are interior construction, packing box and bearing details, and performance and blower shaft horsepower curves.

41—Dual Control System

Clark Bros. Company, Inc., Olean, N. Y. has reprinted J. M. Sappington's article "Dual Control System Synchronizes Compressor Speed - Intake Gas Pressure" which appeared last April in PETROLEUM REFINER. Sappington is superintendent of the Winnie, Texas plant of Absorption Plant, Inc.

Here are the "PRINCO" Thermometers for those really TOUGH service conditions. Cases are extra heavy castings, specially designed for maximum strength. Frames and cases are BOLTED together. Glasses are extra heavy plate to withstand shock and vibration. Cases completely sealed against fumes and moisture. Built-in dehydrator prevents fogging or condensation. Extra protection for mercury tubes assures maximum service life.

Available in all standard case angles and temperature ranges, or to your specification. RED READING MERCURY or Canary-backed tubing optional. Wide selection of "Princo" Separable Sockets in a variety of materials. Write for full information and prices. Ask for Bulletin "E".

NEW! **PRINCO "Easy-Setting" DIAL THERMOMETERS**

5 Standard Scale Ranges
"Easy-Access" Cases for Quick Setting
Simplified Dial and Pointer Adjustment
Rugged, Accurate, Bimetallic Unit

Write for prices and additional information.

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In Canada: PEACOCK BROTHERS Limited

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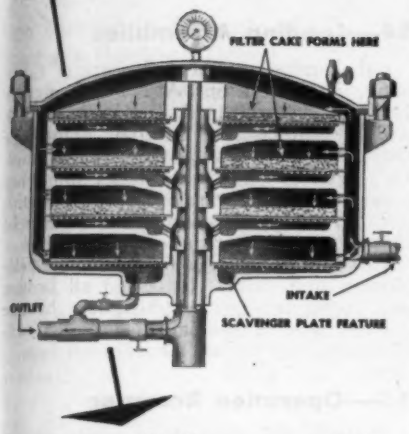

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The Sparkler Horizontal Plate principle insures a firm filter cake with no slipping or cracking. Flow is always with gravity. Filter media is floated into position to form a uniform filter cake of unexcelled stability. The patented scavenger plate (virtually an auxiliary filter) permits complete batch filtration, leaving no unfiltered hold over.

Sparkler filters are equally efficient for continuous or intermittent operation under a wide range of temperature, pressure, and viscosity conditions. Filters are engineered for built-in production line installations or are available in portable units. Standard models made in capacities of 50 to 5000 G.P.H.

We invite correspondence on your problems. You will receive the advice of engineers with a quarter of a century of experience in a specific field.

SPARKLER MANUFACTURING CO.
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42—Stainless Steel Castings

Stainless steel castings from 1 ounce to 2 tons, how they are made and how they are used is the story told in a booklet "A Quarter Century of Pioneering in Stainless Steel Castings" issued by The Cooper Alloy Foundry Company, Department PR, Hillside, N. J. Tables are devoted to analyses, characteristics and applications of standard alloys and also their corrosive and heat uses.

43—Stainless Steel Flasks

Unbreakable and explosion-proof vacuum dewar flasks made of stainless steels and with a built-in radiation shield are described in a leaflet issued by Scientific Glass Apparatus Company, Inc., 49 Ackerman Street, Bloomfield, N. J. New construction features are set forth and new sizes listed.

44—Voltage Control

Electric Machinery Manufacturing Company, Minneapolis, Minn., has issued Bulletin 3100-PRD-192 describing its magnetic "Hi-Fuse" control for motor operation and protection. Cut-away photographs, charts and sketches are included for these devices engineered to meet requirements for motor starting, for motor under-voltage and overload protection, and for high-voltage disconnect.

45—Air Recovery

W. B. Connor Engineering Corporation, 114 East 32nd Street, New York, 16, has published an engineering and application data book (105-A) on its Type "H" "Dorex" air recovery equipment which is used in the "recovery" or conversion of stale, vitiated, or contaminated air to its original freshness, either in order to conserve it for re-use or to correct exhaust and intake odor nuisances. The equipment consists essentially of hollow-walled cylindrical canisters containing specially processed activated carbon and a manifold or supporting plate in which the canisters are secured.

46—Gas Mask Equipment

Acme Protection Equipment Company, 3037 West Lake Street, Chicago 12, is releasing a series of illustrated bulletins presenting a wide range of protection equipment for use in hazardous gas and smoke atmospheres. Information also is provided in table form showing a long list of gases together with equipment recommended.

47—Mechanical Atomizer

The Engineer Company, 75 West Street, New York 6, has issued Drawing 2223-B with notes, describing its mechical atomizer with control for 8-to-1 capacity range within the atomizer head. No returning or recirculating of hot oil is said to be required and high pressure oil is used over the entire load range.

48—Water Treatment

Bulletin G-1 of Liquid Conditioning Corporation, 114 East Price St., Linden, N. J., lists recent developments in the design of water treatment and liquid-conditioning equipment. Included are

. . . YES, you would think we were magicians. Our magic is composed of our engineering know-how *plus* our personnel initiative *plus* our own shops which fabricate standard and special equipment. This formula makes it possible for us to effect short schedules for our clients.

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MECHANICAL ENGINEERING
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FABRICATION
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INITIAL OPERATION

PROJECT
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115 FULTON STREET
NEW YORK 7, N.Y.

tables listing various kinds of gaseous and solid impurities, showing the effects, limits of tolerance etc., and also a comparison chart showing chemical results produced by various water treatment methods.

49—Regulators, Etc.

O. C. Keckley Company, 400 West Madison Street, Chicago 6, is offering a bulletin describing precision pressure regulators, temperature regulators, and combination pressure and temperature regulators, Self-Cleaning Strainers, and Steam Traps. Engineering tables and specifications are included.

50—Stainless Steel Piping, Fittings

Tables showing properties of the most commonly used types of stainless steel, are featured in a bulletin issued by John B. Astwell & Co., 90 West Broadway, New York 7. Discussed is the application of stainless steel pipe, tubing and fittings in various industries where different types of corrosion are encountered. Also tabulated are the dimensions, including wall thicknesses, of Astell light wall, standard and extra heavy grades of pipe which can be supplied in any size for use with Astell fittings.

51—Multiport Drainers

"Multiport" drainers, described in Bulletin 4340 just announced by Cochrane Corporation, 17th St. and Allegheny Ave., Philadelphia 32, are designed for continuously removing large quantities of condensate from evaporators, heaters, separators, coils or steam lines. Condensate is discharged through multiple port openings of a float-controlled rotary valve. Continuous discharge is maintained at a rate determined by the height of condensate in the body. For draining steam separators and similar applications where condensate flow is only a small percentage of steam flow a single port design has been added in the two smaller sizes. The standard drainer is made in five sizes with connections ranging from 1¼ to 5 inches in diameter for pressures to 200 psi. The external cage drainer is made in five sizes with connections ranging from 1½ to 6 inches in diameter for pressures to 300 psi.

For free copies of manufacturers' literature or more information about new equipment featured in Petroleum Refiner, simply identify the product by name and number.

52—Fuel Dispenser

A bulletin describing structural advantages and safety features of newly designed 300-gallon elevated "PX" fuel and lubricant dispensers and improved skid tanks, has been issued by Apex Equipment Corporation, 624 South Michigan Ave., Chicago 5.

Specifications and erection instructions are given for the packaged, all-welded, gravity-feed units, which are delivered in six parts that require only 24 bolted field connections. Also described is a triple-action, self-closing emergency safety valve.

53—Condensate Return System

Cochrane Corporation, 17th Street and Allegheny Avenue, Philadelphia 32, has published an expanded bulletin on high pressure condensate return systems. New technical material on turbulent vs. laminar steam flow and the effects of air and condensate films on heat exchange characteristics are included. Improved equipment for removal and drainage of condensate is also described for the first time. Ask for Publication 3250.

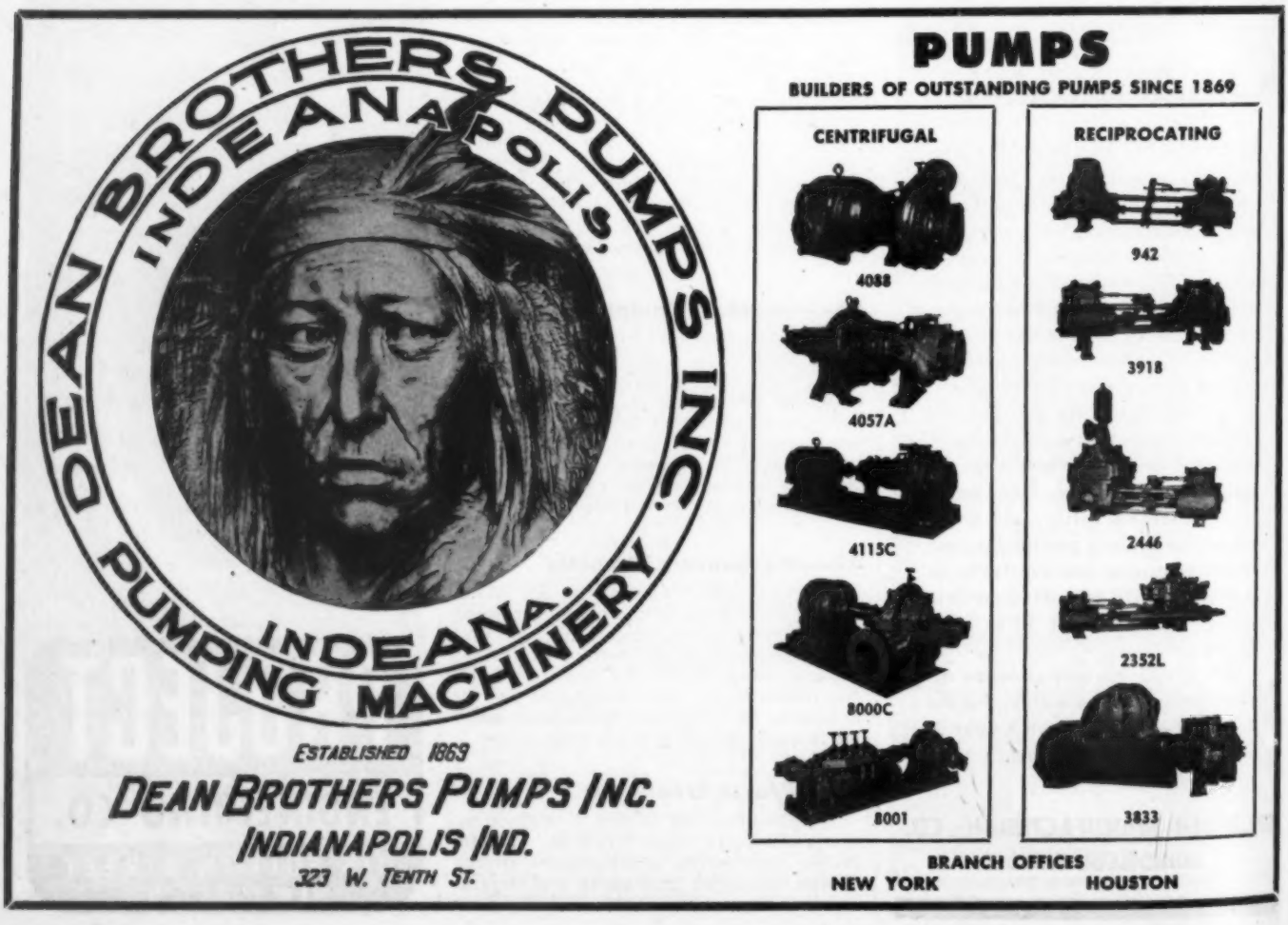
54—Loading Assemblies

Oil Equipment Manufacturing Company, 3100 Vermont Avenue, Louisville 11, Ky., has released Catalog No. 500, which emphasizes flexible dome-type truck and tank car loading assemblies, Oilco swing joints, ball joints and other accessories. The catalog is replete with photographs and schematic drawings of the company's various items.

Introduced is Oilco's perfected self-closing line valves, of tested all bronze and equipped with phosphar bronze springs and a special disc which will withstand up to 300 degrees oil heat.

55—Operation Recorder

Bulletin 247, describing operation recorders is now available, the Esterline-Angus Company, Inc., Box 596, Indianapolis, Ind., has announced. The operation recorder is a multi-pen instrument which can record the time of occurrence, sequence of operation, and duration of as many as 20 events on one chart. Applications include flow of liquids in pipes, emptying and filling of tanks, position of valves, etc.



About EQUIPMENT SERVICE SUPPLIERS

Lee Placed in Charge of Porter's Houston Office

H. K. Porter Company, Inc. of Pittsburgh, has appointed William Lee, district manager of its new Houston sales office. Lee who has been associated with the oil industry for the past 13 years, will handle the complete line of Porter equipment for the processing and refining industries and will also handle the field specialties.

Lee

Evans Joins Branson

Dwight J. Evans has joined Branson Instruments, Inc. Danbury, Conn. as chief engineer, in charge of the development of new instruments for the petroleum trade.

After graduating from Kansas University with a Bachelor of Science in Electrical Engineering in 1938, Evans was connected with the engineering department of Phillips Petroleum Company, Bartlesville, Okla. from 1939 to 1948. He served in the Navy as aviation radar officer for three years, on military leave from Phillips.

Ethyl Announces Changes

Ethyl Corporation has announced changes in the management of its gasoline testing laboratories as follows: William C. Ludt, branch manager of gasoline testing in Tulsa, laboratory to branch manager of the laboratory in North Kansas City, Mo., to succeed Harold L. McIntire, resigned; Charles A. Rankin, branch manager of gasoline testing in San Bernadino, Calif., to branch manager at Tulsa, to succeed Ludt; Robert W. Adams, laboratory technician promoted to branch manager of gasoline testing in San Bernadino, to succeed Rankin.

In addition to Tulsa, San Bernadino, and North Kansas City, Ethyl Corporation maintains gasoline testing laboratories in Baton Rouge, La., and Yonkers, N. Y. As a service to oil refiners,

Guesswork Eliminator

THE NEW DFC "RATIOMATIC" OIL BURNER

Maintains (without variation) any set rates between oil and air over its entire range.

FEATURES:

- Single lever automatic control.
- Control of CO_2 adjustable within 2%.
- Operates on low pressure oil and air.
- 10 to 1 Turndown Ratio.

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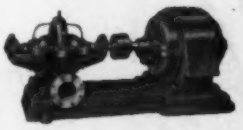
PEERLESS HORIZONTAL CENTRIFUGAL PUMPS



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Single Stage Pump

FOR CONTINUOUS DUTY GENERAL SERVICE PUMPING PEERLESS TYPE A PUMPS

Capacities: Up to 70,000 G.P.M.
Heads: Up to 300 feet
1 to 1000 H.P. • All Types Drive

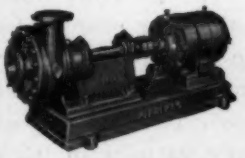
TYPE AF
With Complete Fittings

FOR ADEQUATE FIRE PROTECTION SERVICE TYPE AF FIRE PUMPS*

Underwriters' Approved

Capacities: 500 to 2000 G. P. M.
Pressure: Up to 150 lbs.
20 to 250 H. P.

**Formerly Dayton-Dowd*



TYPE C
Single Stage Pump

FOR ACIDS, CAUSTICS, HOT OILS, CHEMICALS & PROCESS SERVICES...
PEERLESS TYPE C PUMPS

Capacities: 10 to 1200 G. P. M.
Heads up to 100 lbs. • 1 to 60 H.P.

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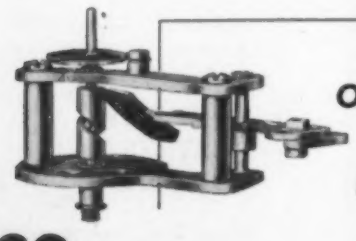
Factories: Los Angeles 31, Calif.; Quincy, Ill.; Indianapolis, Ind. District Offices: New York 5, 37 Wall Street; Chicago 40, 4554 No. Broadway; Atlanta Office: Rutland Bldg., Decatur, Ga.; Dallas 1, Texas; Fresno, Calif.; Los Angeles 31, Calif.

● A vacuum gage indicates the difference between atmospheric pressure (zero on the dial) and absolute zero. Like any Bourdon tube pressure gage, a vacuum gage is really a differential gage. Variations in atmospheric pressure may cause a serious error in the gage reading. This is true particularly when the gage is in a room pressurized by forced draft. A difference of 1 inch in room pressure results in a gage error of about 3½%.

The new HELICOID Vacuum Gage is made so that corrections for changes from 29 to 31 inches in barometric pressure may easily be made by means of the external micrometer pointer adjuster—exclusive with Helicoid. The dial is graduated from 0 to 30 inches mercury vacuum in 0.2 inch subdivisions. It reads counter-clockwise to avoid any confusion with pressure gages, and is designed for greatest ease in reading.

That is another exclusive Helicoid Gage feature.



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TRADE MARK

the company's five testing laboratories regularly test and analyze gasoline samples from all parts of the country and provide consulting facilities on knock testing and related problems.

Fred Ruebelmann Appointed Penberthy Sales Manager

Since Penberthy Injector Company became a subsidiary of the Buffalo Bolt Company, the products, policies and personnel have continued unchanged with one important exception. Fred E. Ruebelmann was appointed general sales manager to succeed A. D. Thomas, resigned. Ruebelmann has been a member of the Penberthy organization for more than 36 years, most recently as supervisor of sales.

Ruebelmann

Student Capacity Doubled at Brown Instrument School

Student capacity, space, and equipment facilities at the Brown School of Instrumentation in Philadelphia has been doubled through the leasing of new quarters. The school now in its 14th year, is conducted for training both customer and Brown Instrument Company service and maintenance engineers.

The new quarters will permit expansion of the school faculty under the direction of M. L. Ladden, according to L. M. Morley, vice president of Minneapolis-Honeywell Regulator Company and its Brown division.

The school will extend its courses in processing control and will add classes in specialized subjects. Free tuition is provided for customer students.

Howell at West Coast Post

Charles M. Howell has been appointed petroleum sales engineer for the West Coast by Detroit Diesel Engine Division, General Motors Corporation. He will have local headquarters in California but will operate out of the companys' petroleum industry sales office, at Tulsa, under the direction of Arch F. Campbell.

Howell came to Detroit Diesel in January directly from the Bethlehem Supply Company of California.

Howell

Linde Products to Expand Its Facilities at Billings

An oxygen filling station and acetylene-producing plant will be built near Billings, Mont., by The Linde Air Products Company, Unit of Union Carbide and Carbon Corporation.

The new facilities will provide industries in the area with a supply of two of the company's main products, Linde oxygen, and Prest-O-Lite acetylene. It is expected that the new facilities will be in operation shortly.

Keating Heads A. O. Smith Southwest Office, Houston

L. M. Keating, assistant to the president in charge of distribution of A. O. Smith Corporation, has been appointed executive administrator of the company's Southwest district office in Houston. He replaces B. F. Bart, recently deceased.

Keating

Keating, a native Texan, was graduated from the engineering school of Washington University. His business experience includes engineering, marketing, and administration, five years having been with the McGraw-Hill Publishing Company. Keating's first assignment with A. O. Smith was to inaugurate a program of product diversification and a later assignment was to develop a national market for welding electrodes and equipment. During the war he supervised the design and construction of the company's Houston electrode plant, which has since been converted to the manufacture of liquid gas systems, pressure vessels, and vertical turbine pumps. Since 1937 he has worked with both manufacturing and sales executives in producing and distributing company products for the petroleum industry. For the past three years, he has served as executive in charge of distribution on the company's management staff.

Ansul Appoints Covert

Ansul Chemical Company, Marinette, Wis., has appointed Kenneth B. Covert to the position of sales manager of its fire extinguisher division. He has been in charge of refrigerant sales at the company's eastern office in Philadelphia since 1945. Previously he was with Macklanburg Supply Company, Oklahoma City, and Revere Brass & Copper Company, Chicago.

Covert

Field Engineers Assigned By Bailey Meter Company

Bailey Meter Company, Cleveland, Ohio, in expanding its staff of field engineers has made assignments as follows: H. E. Bitler to Buffalo, M. J. Bonnette and R. W. Hooper to the Houston area, J. W. Cunningham to Atlanta, H. G. Hilston to Cleveland district office, J. O. Meals to St. Louis, and L. W. Meier to Los Angeles.

J. R. Brennan, J. A. Hays and R. P. Marche are remaining at the Cleveland office where they will be attached to the engineering, research and sales departments respectively.

These men, all recent engineering graduates, have just completed the Bailey Meter Company cadet engineering course. The company also announced two new classes for engineering graduates: one which opened in February and one which is scheduled for July.

- A good strainer protects your equipment and pipelines. The most important part of a strainer is the screen.
- *Screens* used in Yarway Strainers are made of high-grade monel woven wire having small triangular openings fine enough for required stoppage of dirt, rust and scale, but having ample free area for passage of clean condensate, oil or other fluids. Mechanical strength is greater than ordinary mesh. Perforated screens if desired.
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YARWAY STRAINERS
"Police the Pipelines"

Rippie Joins Lukens Steel

Dr. Charles W. Rippie has joined Lukens Steel Company, Coatesville, Pa. as a member of its sales development staff and will specialize in Lukens nickel-, stainless-, Monel-, and Inconel-clad Steels, with partcular emphasis on their applications in the petroleum industry.

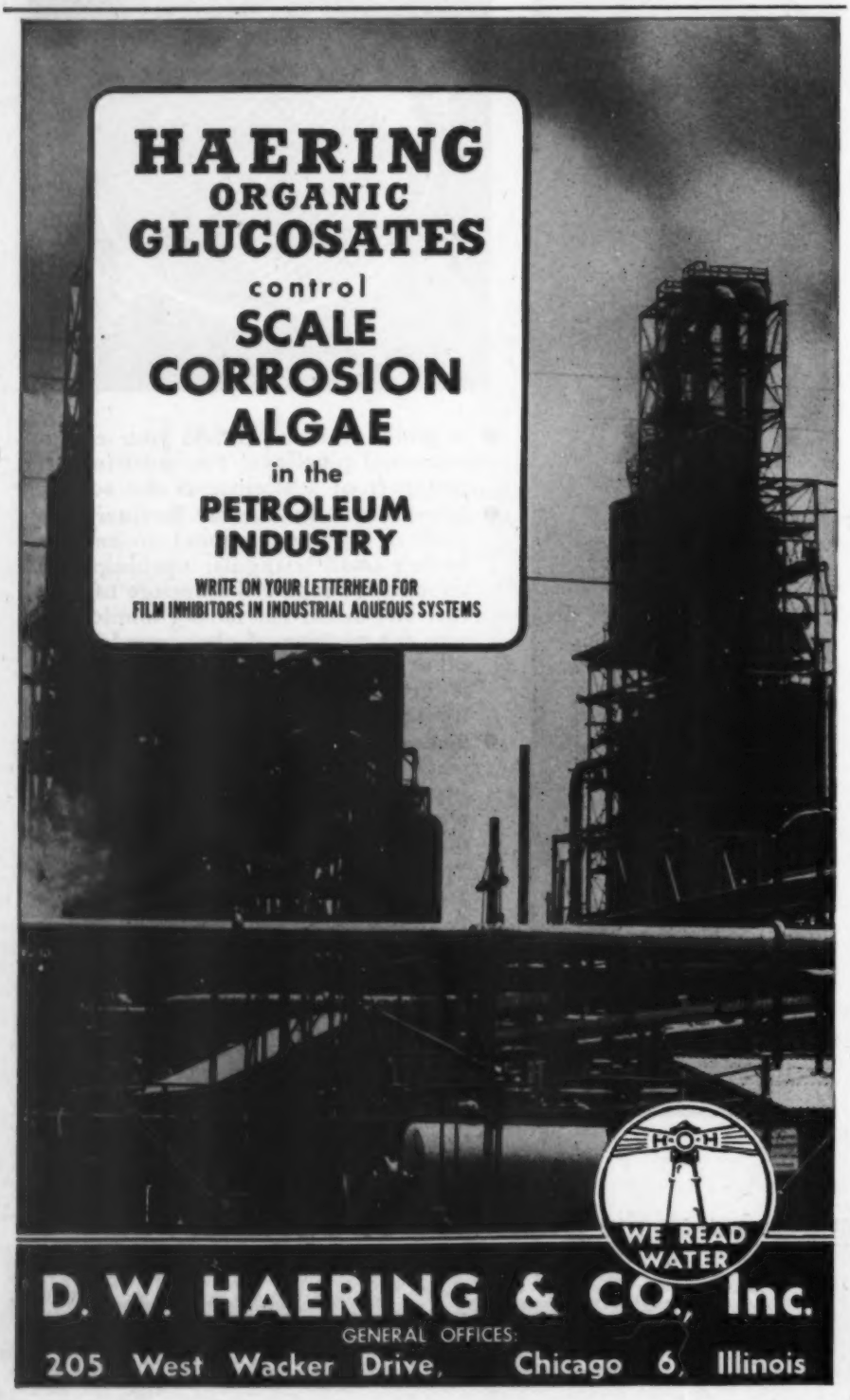
Rippie

Previously, Dr. Rippie had been head of the technical service of the research and development division of Merck & Company, Rahway, N. J. He also was formerly associated with Shell Oil Company, Inc., with Solvay Sales Corporation, where he was petroleum technologist; and with Monsanto Chemical Company.

Wilson to Head Brown Sales in Eastern Area

O. B. Wilson has been named industrial manager for the eastern sales region of the Brown Instrument division of Minneapolis-Honeywell Regulator Company, thus making him regional manager of the territory extending from northern Virginia, through Eastern Pennsylvania, to the Canadian border, west to Rochester, N. Y., and in New England.

Wilson has been with the Honeywell-Brown organization for the past 25 years.



Rear Admiral Hundt Named National Foam New Manager

Rear Admiral L. T. Hundt, USN (Ret.) has been named general manager of National Foam Company's manufacturing plant at West Chester, Pa., and E. P. (Pat) Hughes has been named assistant to the president of National Foam.

Hundt

Admiral Hundt retired from the U. S. Navy last year after 30 years' service. He was chief of staff of the Naval Air Material Center, Philadelphia, from 1946 to July, 1947, when he was retired, at his request, and was advanced to the rank of rear admiral on the retired list.

Admiral Hundt will be in full charge of all manufacturing facilities of National Foam. Hughes will continue to concentrate on sales.

Gregory Succeeds Chambers With Lockett Organization

William Bres Gregory, who has been associated with A. M. Lockett & Company, Ltd., since 1926, and assistant manager of their Dallas office since 1940, has been appointed manager of the Dallas district office, to succeed Henry E. Chambers, Jr., who died March 11. Chambers had been connected with Lockett since 1912.

Gregory joined the Lockett staff in 1926 and was assigned to Dallas in a sales and engineering capacity.

Announcement was also made that Frank P. Elizardi had been named assistant to the vice president of A. M. Lockett & Company, Ltd., with headquarters in Dallas. His work, however, will take him all over Texas.

The Lockett organization has established branch offices in Houston, Galveston, Beaumont, Corpus Christi, and Amarillo, Texas; and, in Shreveport, La.

Anne Named by Elliott

John W. Anne has been appointed sales manager of the Jeannette plant of Elliott Company. He came with Elliott in 1925 immediately after graduating as a mechanical engineer from Penn State. After completing his apprentice course, he was assigned to the condenser department. He became manager of the heat transfer and turbine sales department five years ago.

Walter C. Clapp Dies

Walter C. Clapp, 48, manager of the product sales department of the Manhattan Rubber Division, Raybestos-Manhattan, Inc., Passaic, N. J., died March 6. He had been associated with Manhattan Rubber for 34 years, starting as an office boy.

Berry Moves Office

Effective March 18, Berry Asphalt Company removed its main office from 120 South LaSalle Street, Chicago, to Magnolia, Ark. R. W. Francis and Jim Van Pelt will continue to represent the company from the Chicago office.

Homestead Valve Expands Facilities

Climax of a $400,000 postwar expansion program was marked when Homestead Valve Manufacturing Company formally began production of "Hypressure Jenny" steam cleaners in its new Coraopolis plant, pictured above. Although the youngest branch of the 56 year old company, the "Hypressure Jenny" division outgrew all available space in the main plant. Modern in every detail of design and construction, the new plant has more than a thousand feet of conveyors as well as a motorized assembly line. Incoming materials flow from the freight cars and trucks right into the production and assembly lines with a minimum of handling. This new 40,000-square-foot plant has a rated capacity of 5000 "Hypressure Jenny" steam cleaners a month.

Teeling Appointed Manager

R. F. Teeling of Raybestos-Manhattan, Inc., Manhattan Rubber Division, Passaic, N. J., has been appointed manager of the local sales branch formerly known as "New Jersey Sales," and which will now function as "North Jersey Branch."

Teeling, who has been with the company for 36 years, was assistant manager under Grannell E. Knox, who was manager of the New Jersey Sales Branch for 34 years until his death in January.

GE Promotes Harris

Charles H. Harris has been named manager of the new Decatur, Ill., works of General Electric's plastics division. He first came with General Electric in 1930, shortly after his graduation from Washburn College.

Girdler Names Kennedy

Jack K. Kennedy has been appointed a sales representative of the Votator Division of the Girdler Corporation, Louisville, Ky., and assigned to the staff of Sam N. Welsh, Eastern sales manager, 150 Broadway, New York.

Kennedy, a mechanical engineer graduate of Rose Polytechnic Institute, joined the Votator Division in 1946 after serving in the Army as an officer.

Fabricated Units by Multi-Metal

This grease strainer, shown with a set of inner cartridges, is an outstanding example of Multi-Metal's ability in the design and fabrication of wire and filter cloth units. Units like these have provided industry with the answers to more efficient, less expensive processing operations. Specially fabricated units can be promptly and economically custom-built to your requirements.

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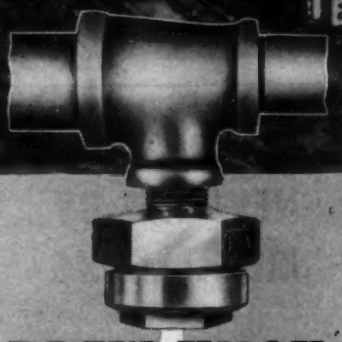

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Rockwell Plant Managers Meet

Plant managers from the 16 divisions and subsidiaries of Rockwell Manufacturing Company recently held their winter meeting at the home offices in Pittsburgh. The gathering permitted interchange of operating experiences in production, engineering, research, accounting, sales, advertising and labor relations at an executive level on a confidential basis and with an unusual lack of restraint, the company explained. Shown above, left to right, are those who attended the meeting, all being from Rockwell manufacturing Company, Pittsburgh, unless otherwise designated: First row: W. S. Potter, W. H. Newbaker, A. J. Kerr, W. F. Rockwell, W. F. Rockwell, Jr., E. W. Meyers, C. A. Wiken. Second row: H. C. Stuckeman, Delta Manufacturing Co.; J. E. Ashman, H. S. Rockwell, Hopewell Division; A. E. McIntyre, B. C. Trueblood, Arcade Manufacturing; W. R. Brown, Arcade Manufacturing; A. R. Whittaker, National Motor; C. H. Daugherty, Monessen. Third row: R. G. Caouette; N. J. Kenney, Ohmer Corporation; A. E. Higgins, R. I. C.; L. W. Heavener, Macnick Division; R. C. DuBruoq, Crescent Machine; E. F. Foubert. Top row: H. C. Greene; W. F. Crawford, Edward Valves; Floyd Snodgrass, Nordstrom Valve, Oakland; Wm. A. Marsteller; C. S. Bygate; L. A. Dixon, Jr., Pittsburgh-DuBois.

American Brake Shoe Lists Official Changes

Fred P. Biggs has been appointed first vice-president of Brake Shoe & Castings Division of American Brake Shoe Company. In addition to his new duties, he will continue as Vice-President in Charge of Sales of the Brake Shoe & Castings, and Southern Wheel Divisions.

Other Brake Shoe appointments announced last month include: Roger W. Batchelder appointed assistant to the president of the National Bearing Division of American Brake Shoe Company and William H. Old appointed general purchasing agent for American Brake Shoe. Batchelder will be located at the division headquarters in St. Louis, whereas Old will continue to be located in New York.

Ritschard Transferred

Bailey Meter Company, Cleveland, Ohio, has transferred R. B. Ritschard from its Cincinnati office to its Cleveland district office. Ritschard has been with Bailey Meter Company since 1936.

FIRE PUMPS

36—New HALE High Pressure Centrifugal Pumps—500 GPM @ 346 Ft. Hd.—4" Suction and Two 2½" Discharge Connections equipped each with Quick Opening Valve. Direct connected to Chrysler Ind. 7, 6 Cylinder Gasoline Engine. Units complete with starter and generator — Steel Skid Mounted — Boxed for Export.

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